The effectiveness of antiscalants when used on multi-stage flash distillers.

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REFERENCE

The Effectiveness of Antiscalants when used on Multi-Stage Flash Distillers

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ABSTRACT

This project is concerned with the testing of antiscalants (scale inhibitors) used to improve the performance of Multi-Stage Flash (MSF) distillers in Kuwait.

This study describes and compares the thermal schemes, thermal efficiency and thermal performances ratios of the distillers when treated with three different antiscalants over a testing period of 6 months. The plants used for the tests, the Doha West and AZ-Zour power stations, which have a capacity of producing 7.2 million imperial gallons of potable water per day are described.

The evaluation tests revealed that all the examined antiscalants were successful in inhibiting alkaline scale formation and improved the plant performance at top brine temperatures ranging between 90°C and 110°C with dosing rates of 1.5 and 3.0 ppm. The results of the products (Belgard EV 2030; Sokolan PM 10i and NCP DS 3000) show an insignificant difference in the values of overall heat transfer coefficient and fouling factor of the heat gain section.

Relationships between the parameters controlling the distillate water product to other operating and design parameters are presented. These parameters include plant performances rates, flow rate of re-circulating brine and top brine temperature.

The evaluation tests were based on the measured and calculated thermal parameters, distillate water production rate, Gain Output Ratio (GOR), Performance Ratio (PR), Top Brine Temperature (TBT), antiscalant type, and dosing rate.

The antiscalant dosing rate was calculated for each day together with the overall heat transfer coefficients and fouling factors for the brine heater and the heat recovery sections. Recorded data and calculated parameters have been tabulated chronologically. Based on the recorded and calculated data, graphical plots were generated to compare each critical parameter against the elapsed time for the entire test periods.

Each of the three antiscalants tested resulted in the successful operation of the MSF distillers at low antiscalant dosing rates. The performance testing revealed that the best scale control test was able to operate at a high temperature (105-107°C) and low additive dosing rate.

The use of improved antiscalants gives the possibility of increasing water production as reduced scaling will allow a number of currently operating MSF plants to increase their top brine temperatures.

PREFACE

This work presented in this thesis was carried out between January 2007 and October 2011 at Sheffield Hallam University, Department of Engineering and Mathematics / Faculty of Arts, Computing, Engineering and Sciences and at Doha West Power Station.

The work described is original, excepted where reference has been made to others, and no part of it has been submitted for an award at any other College or University.

During the course of work, papers presented at the following conferences:

- 10th Middle East Corrosion Conference & Exhibition, 7-10 March 2004, Gulf International Convention Centre, Bahrain.

- EuroMed 2006 conference on Desalination Strategies in South Mediterranean Countries: Cooperation between Mediterranean Countries of Europe and the Southern Rim of the Mediterranean, European Desalination Society and the University of Montpellier, France, 21-25 May 2006.

- Conference on Desalination and the Environment, European Desalination Society and Centre for Research and Technology Hellas (CERTH), Sani Resort, Halkidiki, Greece, April 22-25, 2007.

- The 1st Maghreb Conference on Desalination and Water Treatment, Tunisian Desalination Association, Hammamet, Tunisia, December 7-10, 2007.

- EuroMed 2008 Desalination, Cooperation among Mediterranean Countries of Europe and the MENA Region, November 9-13, 2008, Dead Sea, Jordan.

- International Desalination Association, Perth Convention and Exhibition Centre (PCEC), Perth, Western Australia September 4-9, 2011.

- EuroMed 2012 Desalination for the environment clean water and energy conference and exhibition, April 22-26, 2012, Barcelona, Spain.

The following papers relating to this work have been published, the copies of which may be found in the appended CD:

- Towards More Economical and More Reliable Desalination Plants: A New Innovative Antiscalant Control Management in MSF Plants.

- Fleat Transfer measurement as a criterion for performance evaluation of scale inhibition in MSF plants in Kuwait.

- The release of C02 in MSF distillers and its use for the re-carbonation plant: a case study.

- Impacts of different antiscalants dosing rates and their thermal performance in multi-stage flash (MSF) distiller in Kuwait.

- Aspects of operating experience of different antiscalants in multi-stage flash (MSF) distillers.

- The effectiveness of antiscalants when used on multi-stage Flash distillers in Kuwait.

  Chemical analysis of distilled water - A case study.

ACKNOWLEDGEMENTS

In the name of ALLAH, most Gracious, most Merciful, First and Foremost, I would thank ALLAH, the Almighty, the ALL-Hearing, and the most Bountiful for his grace giving me the opportunity to complete this thesis successfully.

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I also appreciate the extended excellent technical support and facilitation provided by Kuwait's Doha West Power Station *staff* with special thanks to *them* for their hard works, co-operation and assistances.

Sincere gratitude and loving thanks goes to my *husband for* his unconditional love, encouragement and support. This thesis would not have been possible without him. Gratefulness should also go to my *family,* especially, my *mother, father, sisters* and *brothers* for their *love,* encouragement and faith in me.

Finally, most thanks go to ALLAH for giving me the chance, strength and courage to perform and complete this work 'AL-HAMD-LELLAH'.

GLOSSARY

Alkalinity : The quantitative capacity of aqueous media to react with hydrogen ions.

Antiscalant Chemical : A chemical inhibitor which prevent alkaline scale components from precipitating on the heat exchanger tubes.

Brackish water : Water that contains dissolved matter at an approximate concentration range from 1000 to 30000 mg//.

Brine : Concentrated seawater with a dissolved solid content greater than the dissolved solid content of seawater.

Brine Blow down : The most concentrated brine that leaves the evaporator final stage flash chamber, which includes the dissolved solids in the evaporator, feed that are concentrated into the brine blow down.

Brine Heater : The heat input section of a multi stage flash evaporator where feed water is heated to the process temperature.

Concentration Ratio : The ratio of total dissolved solids in recycled brine to that of seawater.

Condensate : Liquefied heating steam on the heat transfer tube surfaces of the brine heater.

Conductivity : The reciprocal of the electrical resistance, a measurement of the electrical conductivity of water is a guide to the presence of dissolved solids or gases. Seawater and brine have a relatively high conductivity whereas condensate and distillate have a very low conductivity.

Demister : Layers of knitted stainless steel wire mesh pads to intercept salt water droplets carried over with the generated vapour by flash evaporation.

Desalination : The process of removing dissolved salts from seawater.

De-superheater : Spray nozzle installed in the steam line entering the brine heater. The condensate from the brine heater is sprayed and atomized to lower the heating steam temperature and saturate it so that it will more readily condensed on the

heat exchanger tubes of the brine heater.

Distillate — Condensed vapour accumulated on the heat transfer tube surfaces of the evaporator condenser, that is, product water.

Distillate Tray — A bed plate situated under each evaporator stage condenser tube bundle to collect the distillate generated at each stage of evaporation.

Distillate Trough — A flow channel to transfer the distillate collected at each stage of evaporation to the next adjacent stage.

Evaporation — The process in which water is converted to a vapour that can be condensed.

Flash chamber — The internal part of the evaporator beneath the distillate tray and demister. The brine passes through a series of flash chambers so that some brine will flash in each chamber.

Flashing — A method of vaporizing heated brine by a pressure reduction. Flashing will occur when the hot brine is subjected to a pressure below its vapour pressure.

Flashing Range — The difference between the maximum brine temperature and the bottom temperature in the last stage.

Gain Output Ratio (GOR) : A measure of distillate performance which represents the ratio of mass flow of distillate to steam input in heat input section.

Heat Input Section — The heat exchange section of the evaporator in which external heat is added, i.e. the brine heater.

Heat Recovery Section — The heat exchange section of the evaporator in which heat from condensing vapour is recovered by re-circulating brine.

Heat Rejection Section — The heat exchange section of the evaporator in which heat from condensing vapour is removed by the cooling seawater as waste heat.

Make-up — Seawater feed to the evaporator which makes up the loss of recycled brine that is removed from the desalination system as product water and blow down brine.

Non Condensable Gases :	Gases such as carbon dioxide and other constituents of air which do not condense under the given condition. These gases must be removed to prevent them from forming a heat insulation layer on the heat exchanger tubes.
Performance Ratio	The ratio of mass of distillate produced per unit of heat transfered.
pH	The negative logarithm of the hydrogen-ion activity in an aqueous solution, or, the logarithm of the reciprocal of the hydrogen-ion activity.
Product	The fresh water produced in an evaporator, i.e. the distillate.
Reverse Osmosis (RO)	A desalination process using continuously applied high pressure to the feed water, pushing water through a semi permeable membrane.
Seawater	Seawater contains about 3.4% salt and is slightly alkaline, pH 8. It is a good electrolyte and can cause galvanic corrosion and crevice corrosion.
Scale	Salts deposited on heat transfer surfaces which retard the rate of heat transfer. These salts, precipitated from locally supersaturated brine, are classified either as alkaline such as calcium carbonate and magnesium hydroxide, or non-alkaline such as calcium sulphate.
Shell side	The outside of heat exchanger tubes.
Sludge	A water-formed sedimentary deposit.
Stage	The individual chamber of the evaporator.
Tube Side	The inside of heat exchanger tubes, the opposite of shell side.
Vapour	The steam generated from the hot brine in the flash chambers of the evaporator. In general, the word “steam” refers to the water vapour generated in the boiler of the power plant.

NOMENCLATURE

A		Area of heat transfer surface (m2).
Cp		Specific heat at average temperature of inlet and outlet brine heater temperature (kJ/kg K).
FF		Fouling factor (m2K/kW).
TBT		Top Brine Temperature (°C).
T0		Brine heater outlet temperature (°C).
Tj		Brine heater inlet temperature (°C).
Tc		Brine heater condensate temperature (°C).
Ti	:	Re-circulating brine of stage no.1 temperature (°C).
T2i		Re-circulating brine of inlet stage no.21 temperature (°C).
Tvi		Vapour temperature of stage no. 1 (°C).
TV2i		Vapour temperature of stage no.21 (°C).
U		Calculated heat transfer efficiency (kW/m2K).
Uc		Overall heat transfer coefficient at the initial condition (clean value) (kW/m2K).
URec		Heat transfer coefficient of heat recovery section (kW/m2K).
Wbr		Re-circulating brine flow rate (tonnes/h).

ABBREVIATIONS

ABVC	Absorption vapours compression.
AC	Alternating Current.
ADVC	Adsorption vapours compression.
BH	Brine Heater in Desalination Process.
CaCOs	Calcium Carbonate.
C02	Carbon Dioxide.
COD	Content of Dissolved Oxygen.
CVC	Chemical vapour compression,
d	Day.
EDTA	ethylene di-amine-tetra-acetic acid
DB	Deutsche Babcock.
DWPS	Doha West Power Station.
ED	Electro-Dialysis.
EDI	Electro-Deionization.
GCC	Gulf Cooperation Council
GOR	Gained Output Ratio (kg distillate per kg steam).
h	Hour
IDA	International Desalination Association
LP	Low Pressure.
MED	Multi-Effect Distillation.
MEE	Multi-Effect Evaporation.
MEW	Ministry Electricity and Water of the State of Kuwait.

MIG	Million Imperial Gallons.
MIGPD	Million Imperial Gallons per Day.
MSF	Multi-Stage Flash Desalination Technology Process.
MVC	Mechanical vapour compression.
pH	pH value determines whether water is hard or soft, measurement of pH to determine the corrosiveness of the water.
ppm	Parts per million.
PR	Performance ratio.
SEE	Single effect evaporation.
SEE-VC	Single effect evaporation with vapour compression.
S02	Sulphur Dioxide.
SS	Stainless Steel.
TDS	Total Dissolved Solids.
TVC	Thermal vapour compression.
VC	Vapour compression.
WHO	World Health Organization.

TABLE OF CONTENTS

Introduction

Aims, Objectives and Thesis Overview

This thesis is concerned with the thermal performance of Multi-Stage Flash (MSF) desalination plants as affected by the seawater being processed which contains salts that can be deposited as scale and sludge and can retard the heat transfer process resulting in serious loss of efficiency and production, For these reasons various types of chemical antiscalant additives (scale inhibitors) have been developed to control scale formation and this study will evaluate the performance of three antiscalants over an extended trial period. The heat transfer coefficients, fouling factor and Gain Output Ratio (GOR) were taken as parameters to evaluate the plant performance and hence the effectiveness of the antiscalants.

Many countries are dependent on water produced by desalination technologies and comparisons will be made between the different processes in current use. The distillation processes of this study are being continually developed and a literature review will be conducted to highlight the various stages of development and how they have been improved. The nature of scale deposits will be investigated together the chemical processes involved and the mechanisms used by antiscalants to control the build-up of scale.

This study will provide an overview of desalination technologies, particularly Multi-Stage Flash evaporation technology, scale control additives and corrosion control. The use of antiscalant additives is one of the most frequently used methods to control the scale formation on the heat transfer surfaces of MSF distillers.

The aim of this study is to conduct extended trials to determine the performance of using three different antiscalants types on 7.2 MIGPD (million imperial gallons per day) Multi-Stage Flash distillers which have operated satisfactorily over 6 month periods of time. Reduced antiscalant dosing rates were trialled and the performance measured to test the efficacy at the lower dosing rates.

All MSF plants in Kuwait use chemical additive dosing, which are capable of performing at high top brine temperatures, to effectively inhibit scale formation

An important objective will be to develop a monitoring and testing procedure for a practical field trial to provide data that will be used to enable plant performance to be measured in order to compare the three antiscalants under test. The Doha West (DWPS) and the AZ- Zour Power Stations were used as the reference power plants in this study. The study will also indicate the important chemical and operational conditions of the existing distillers.

A summary survey of historical developments and more recent applications of the use of antiscalant chemicals in existing power and water plants in Kuwait and other countries are investigated and outlined in this present thesis. Relationships of the parameters controlling the distillate product to other operating and design parameters are presented. These relationships are used for studying the performance of operating distiller plants.

The contribution made to the body of research knowledge will be to improve on previous performance trials and to leave in place a facility that can be used to provide a test bed to trial new antiscalants and to investigate changes in running conditions that can improve plant performance. The trial will establish whether dosage rates lower than the industry norm is effective in controlling scale formation. This is an important issue as lower dosage rates reduce operational costs but more importantly reduce chemical emissions and the environmental impact they have. One of the antiscalants used in this current study was previously trialled by Ata Y. Abdulgader et.al (1) at the Jeddah IV commercial plant in Saudi Arabia but at a higher dosing rate of 2 ppm.

This research thesis investigates optimization to improve the chemical process at various phases of design, operation, and control of a plant.

To achieved the goal of an efficient MSF process the proposed study aims to:

- Develop an understanding and awareness of technical capabilities of antiscalant chemicals.
- Minimize energy consumption; this means achieving high GOR for the plant, which is defined as the ratio of distillate production rate to the steam input rate.
- Achieve stable operation: by controlling the top brine temperature and the brine levels in the flashing chambers.
- Improve demister pad fouling which would allow the plant to be operated without shutdown of the plant for demister pad acid cleaning.
- Avoid heat exchanger scaling and sludge by adjusting the operational mechanical and chemical parameters: Minimise chemical consumption, thus its cost while maintaining effective plant performance.
- Attain a desirable brine maximum temperature in order to improve plant efficiency.
- Avoid high maintenance and reduced output by not having to remove scale mechanically or by acid cleaning which also reduces the plant life.
- Establish a strategy so that alternative suitably tested antiscalants can be substituted when the suppliers of currently used antiscalants increase their prices

Research setting

The present studies were instigated following recommendations of the Doha West chemical committee for which the author is head. The study is based on plant operation of two 24-stage MSF desalination plants (7.2 MIGPD) which are currently operating in Kuwait.

Currently, there are seven stations in Kuwait for the generation of power and the production of distilled water. When Kuwait began to build the current type of MSF plants in the late fifties, there was a limited choice of simple antiscalant chemical additives.

Recently, a number of antiscalant chemicals have been introduced into the market, that permit the MSF plants to operate at higher top brine temperatures which enable to plant to be more thermally efficient.

A number of these newly developed antiscalant chemicals have been presented to the Ministry of Electricity and Water (MEW) by various manufacturers for trial evaluation. Each chemical undergoes preliminary evaluation based on the data supplied which have been generated by existing MSF plants. A recommendation for the trial of the chemical by MEW is dependent upon the reliable and reasonably effective outcome of this preliminary evaluation.

The MEW stated policy requires that every available antiscalant chemical must be properly tested in one of its plants under its full control to fulfil all the requirements and complying with MEW protocol, in order to determine its performance. Only after the completion of these tests to MEW satisfaction, would the antiscalant chemical be considered on a commercial basis. This includes two important considerations that of safe handling and a non-corrosive nature.

To obtain reliable performance data on the new antiscalant chemicals, the trial should be over an extended period of time, using calibrated instrumentation, so as to quantify the effectiveness of antiscalant chemicals in terms of distillate production, GOR, heat transfer coefficients and the fouling factors of the brine heater and heat recovery sections. Many short trials have been carried out in Kuwait to demonstrate comparable performances. The current trial of three new antiscalant additives was carried out the DWPS and AZ-Zour power station. The antiscants and testing period being:

Antiscalant A: Belgard EV 2030

Testing Station: AZ-Zour, Distiller D5
Distillate Production rate: 7.2 MIGPD (High Operation)
Top Brine Temperature 105°C
Ball cleaning in operation
Starting date: 26/03/2000

End date: 09/09/2000

The trial was started on the first week with a dosing rate 3.5 ppm, which was reduced to 3.0 ppm for four weeks and then, continued for the last 19 weeks at 1.5 ppm

Antiscalant B: Sokolan PM 10i

Testing Station: DWPS, Distiller D5A
Distillate Production rate: 7.2 MIGPD (High Operation)
Top Brine Temperature 105° C
Ball cleaning in operation
Starting date: 14/05/2001
End date: 15/09/2001

The trial was started at the rate of 3.0 ppm and reduced to 1.5 ppm for four weeks and then, continued for the last 14 weeks at 1.5 ppm

Antiscalant C: NCP DS 3000

Testing Station: DWPS, Distiller D8B
Distillate Production rate: 7.2 MIGPD (High Operation)
Top Brine Temperature 105°C
Ball cleaning in operation
Starting date: 28/01/2003
End date: 03/08/2003

The trial was started on the first week with a dosing rate of 4.0 ppm, which was reduced to 3.5 ppm for two weeks and then, reduced to 3.0 ppm for two weeks, reduced to 2.5 ppm for two weeks, reduced to 2.0 ppm for two weeks, reduced to 1.6 ppm for three weeks, then continued for the last 98 days at 1.5 ppm.

The effectiveness of the antiscalant additive was to be judged by monitoring all aspects related to the overall performance of the unit and the impact the additive might have on the operation of the unit and its performance. It was necessary, therefore, to ensure that unit operation was as near as possible to design conditions.

In order to ensure that the performance data collected from the trial of any additive is relevant, it was imperative that such trials should be carried out on a distillation unit which is of modern design, representative capacity and equipped with adequate instrumentation.

Thesis Overview

The dissertation comprises of seven chapters and six supporting appendices:

- The scope of *Chapter 1* is an introduction to desalination technology history and development.

- *Chapter 2* gives an extensive literature survey of scaling, foaming, corrosion, scale control additives, antifoaming additives, and the Kuwaiti experience of scale control.

- *Chapter 3* describes Multi-Stage Flash distillation, scale control methods and operating conditions of MSF plants.

- *Chapter 4* presents a detailed description of the experimental procedure (methodology), equations and sample calculations of overall heat transfer coefficients and fouling factors of the brine heater and heat recovery section, instrumentation selections and calibration procedures, pre and post-trial inspections, data analysis and evaluations, and operational and performance parameters.

Chapter 5 presents an analytical comparison study between the three different studied antiscalants. The effectiveness of the control of the plant during the tests is examined. Variations in the measured parameters are commented upon which include variations in ambient conditions, drift in plant operation and plant malfunctions. The magnitudes of the variations are quantified by the mean, standard deviation and the coefficient of variation. The results of the analysis of scale samples are presented which provide data on the chemical composition of scale deposited within the plant.

The conclusions are given in *Chapter* 6 as well as recommendations for future work.

Finally, *Appendix A* presents chemical and physical properties of water.

Appendix B gives a background of the history of water supply in Kuwait together with details of the installed capacity of MSF plant.

Appendix C includes a description of different types of scale control additives.

Appendix D describes the ball cleaning system

Appendix E illustrates the chemical analysis of the MSF plant procedure.

Appendix F includes instrumentation photographs with their specifications.

Appendix G (CD rom) provides detailed spreadsheet data and analysis for the three antiscalants.

Desalination Processes

1.1 Need for Desalination

Water is an essential commodity for life especially for industrial societies that relies on water for manufacturing, agriculture, domestic and leisure activities. Growing populations and increases in personal demand for water has led the need to exploit existing and new sources. For many locations, with a shortage of natural water supplies, they are now reliant on desalination of the abundant availability of seawater.

Khawaji et al (2) proved numerically, “Available fresh water resources from rivers and groundwater are presently limited and are being increasingly depleted at an alarming rate in many places. The oceans represent the earth’s major water reservoir. About 97% of the earth’s water is seawater while another 2% is locked in icecaps and glaciers. Available fresh water accounts actually for less than 0.5% of the earth’s total water supply. Virtually by removing salt from the unlimited supply of seawater, desalination has emerged as an important source of fresh water”.

Because of huge demands for the requirements of potable water, many countries are now using desalination technologies. Principally, in the Arabian Gulf area this includes countries such as Saudi Arabia, United Arab Emirates, Bahrain, Qatar, Oman and Kuwait.

The International Desalination Association (IDA) inventory in 2004 (3) stated that “over 75 million people worldwide obtain fresh water by desalinating seawater or brackish water”. Even countries with abundant rainfall such as the United Kingdom require desalination plant in some regions to cover periods of low rainfall as it is less expensive than pumping water from distant reservoirs.

The IDA director, Leon Awerbuch (4) reported that as of June 30, 2008, the cumulative contracted capacity of desalination plants around the world was 62.8

million m^3/day and the cumulative installed capacity was 52.3 million m^3/day. This production was from 13,869 desalination plants worldwide.

The World's top 10 producers are given in table 1.1.

No.	Producer Country	Production (m^3/d)	%
1	Saudi Arabia	10,759,693	17
2	UAE	8,428,456	13
3	USA	8,133,415	13
4	Spain	5,249,536	8
5	Kuwait	2,876,625	5
6	Algeria	2,675,958	4
7	China	2,259,741	4
8	Qatar	1,712,886	3
9	Japan	1,493,158	2
10	Australia	1,184,812	2

Table 1.1: Top 10 country producers (4).

At the end of 2002, Multi Stage Flash (MSF) and Reverse Osmosis (RO) accounted for 36.5% and 47.2%, respectively, of the installed brackish and seawater desalination capacity (4). And, "With 97% of available water represented by salty water with the Salinity Level > 35,000 mg/l, the largest possible source of alternative water supply requires and will require desalination" (4), as shown in Figure 1.1.

Water Resources On the Earth

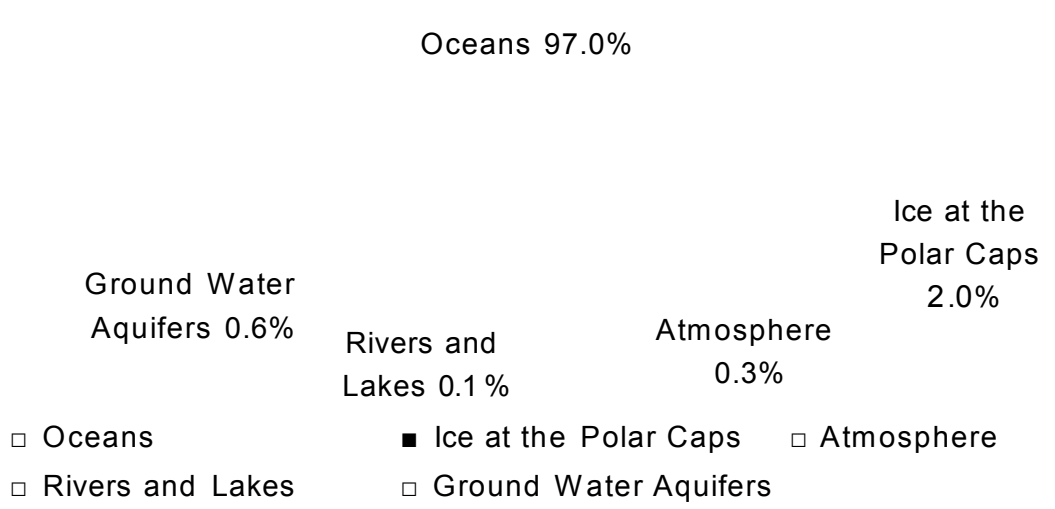


Figure 1.1: Water Resources on the Earth (5).

Desalination technology has developed for over a millennium, very slowly at first and then its development started to accelerate.

Although water is one of the most common of substances and is comprised of only two elements it has extraordinary, anomalous and complex properties. It can act as an acid or a base and has great solvent power for ionic compounds and can readily dissolve many non-ionic covalent compounds. More details of the nature of sea, ground and river waters and their properties are presented in Appendix A.

A. Al-Radif (6) said that “Desalination of seawater is fast becoming a major source of potable water for long-term human survival in many parts of the world. Among the seawater desalination processes, the MSF process produces large scale potable water more successfully than many of the alternative process”.

Also, K. A. AL-Shayji said that “An additional advantage of an MSF plant is the capability of coupling it to a power generation plant as the heat source, thus making the process increasingly important for water and power production”.

1.2 Desalination Technology History

The ancient Greeks are said to have used evaporation from sea water to obtain drinking water. It is claimed that the first desalination plant in America was turning seawater into drinking water at Fort Zachary Taylor in Key West, Florida as early as 1861 (7), but the use of modern technology for desalination probably dates from the beginning of the last century.

Potable water in Kuwait is produced by desalination of seawater in multistage flash desalination (MSF) plants. In 1914 the first desalination plant in Kuwait was commissioned. Figure 1.2 shows a distillation plant which was installed in 1945, and throughout the 1950s and 1960s a large number of desalination plants, mainly thermal, were installed around the world for both irrigation and drinking water supply, mainly by British companies.

The first electro-dialysis (ED) desalination plants were installed by William Boby in 1963 in Libya. Although reverse osmosis had been recognised as a potential technology for desalination in the early 1960s and a few RO desalination plants had been constructed in the late 1960s, it was many years before membranes of sufficient quality and appropriate cost were developed to permit RO to start competing with thermal processes (8).

The historical development of the water supply in Kuwait and details of the installed capacity of MSF plant are given in Appendix B.

Figure 1.2: General view of the first desalination plant that was established in Kuwait, (9).

1.3 Desalination Processes

Tsiourtis (10) explained the desalination process as following: "The desalination process separates nearly salt-free water from sea or brackish water. The desalinated water is recovered for consumption and the salts are concentrated in a stream of water called the brine reject, disposed either to the sea or to a saline aquifer or in evaporation ponds. The process requires energy to operate and can use a number of different technologies. A typical flow diagram of the process with inputs and outflows is shown in Figure 1.3 the commercially available desalination processes are based on thermal or membrane methods".

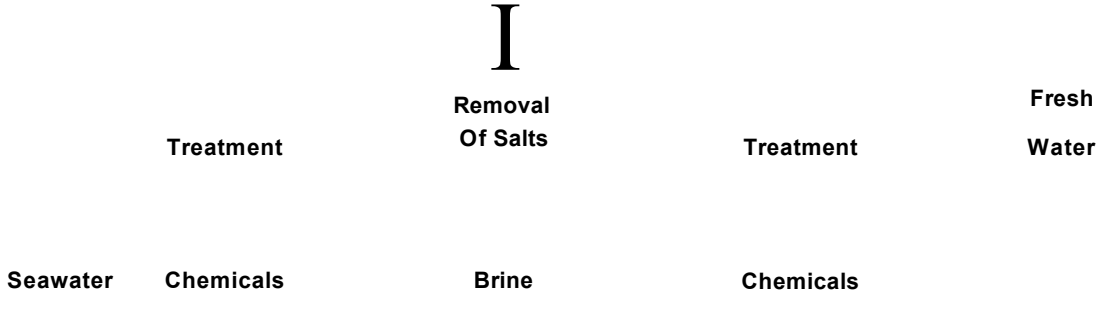


Figure 1.3: Desalination process (10).

1.4 Classification of Desalination Processes

The desalination process can be illustrated as a chemical engineering concept. Seawater fed to the desalination technology with heat energy produces desalinated water and concentrated brine as shown in Figure 1.4.

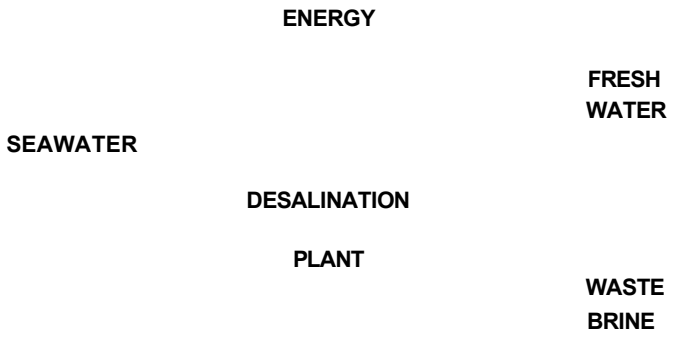


Figure 1.4: Chemical Engineering Technology Concept (8).

H. El-Dessouky (11) has classified the different desalination presses as shown in Figure 1.5.

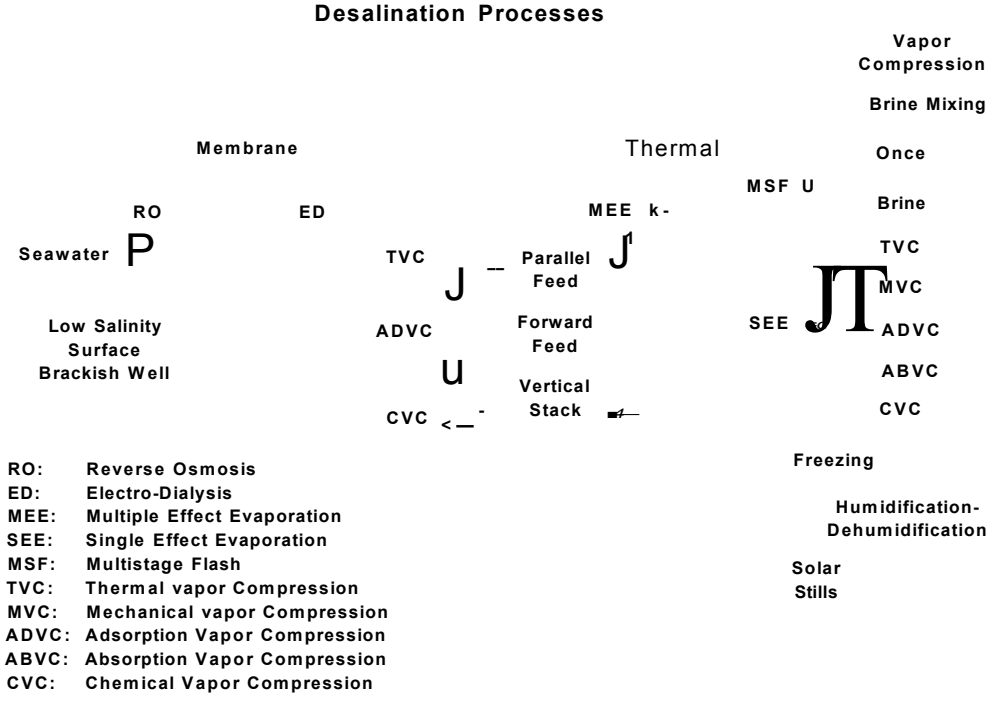


Figure 1.5: Classification of thermal and membrane desalination processes (11).

1.4.1 Thermal Processes

The simplest example of a thermal process is distillation. Saline water is heated to generate steam, which is then condensed to produce non-saline water. Several forms of distillation are used for desalination (8).

- Multi-Stage Flash Evaporation / Distillation (MSF)
- Multipie-Effect Evaporation (MEE) / Distillation (MED) also known as Long-Tube Vertical Distillation (LTV)
- Vapour Compression Distillation (VCD)
- Solar distillation

There are three thermal processes which use thermal energy to freeze and then melt the ice formed in the freezing stage (8).

- Vacuum freezing
- Secondary refrigerant freezing
- Hydrate formation process

Evaporation processes include MSF, MED, Single Effect Evaporation with Vapour Compression SEE-VC, humidification-dehumidification, and a number of methods based on the use of solar energy (12).
Water Desalination Technologies in the Economic and Social Commission for Western Asia (ESCWA) Member Countries report (12) illustrated VC as following: “VC is combined with single or multiple effect desalination processes to improve thermal efficiency. In the VC process, low* temperature vapour formed in the same effect or the preceding evaporation effect is compressed and used to initiate the evaporation process in the first or the same evaporation effect”.

The VC process incorporates component devices that include mechanical compressors, steam-jet ejectors, Thermal Vapour Compression TVC components, absorption/absorption beds, and absorption/absorption columns. Variants of the single effect VC process

include Mechanical Vapour Compression MVC, Absorption Vapour Compression ABVC, Absorption Vapour Compression ADVC, and Chemical Vapour Compression CVC (12).

1.4.2 Membrane Processes

The membranes used by these processes filter dissolve salts by synthetic polymeric membranes by applying a gradient pressure or a potential electrical difference through the membrane layers. The types of membrane process are:

- Reverse Osmosis (RO).
- Electro-dialysis (ED).

Figure 1.6 shows schematics of the main conventional processes (13). In the RO process, high pressure forces fresh water to permeate through a semi-permeable membrane, leaving behind a highly concentrated brine solution.

- Reverse osmosis is becoming the most widely used method for the desalination of brackish and sea waters. RO plants vary from small domestic units for use either in the home or on small ships, to large industrial and municipal units for supplying communities with a potable water supply.

- The RO process is less energy intensive and consequently a favoured option in countries which do not have large oil and gas resources at their disposal.

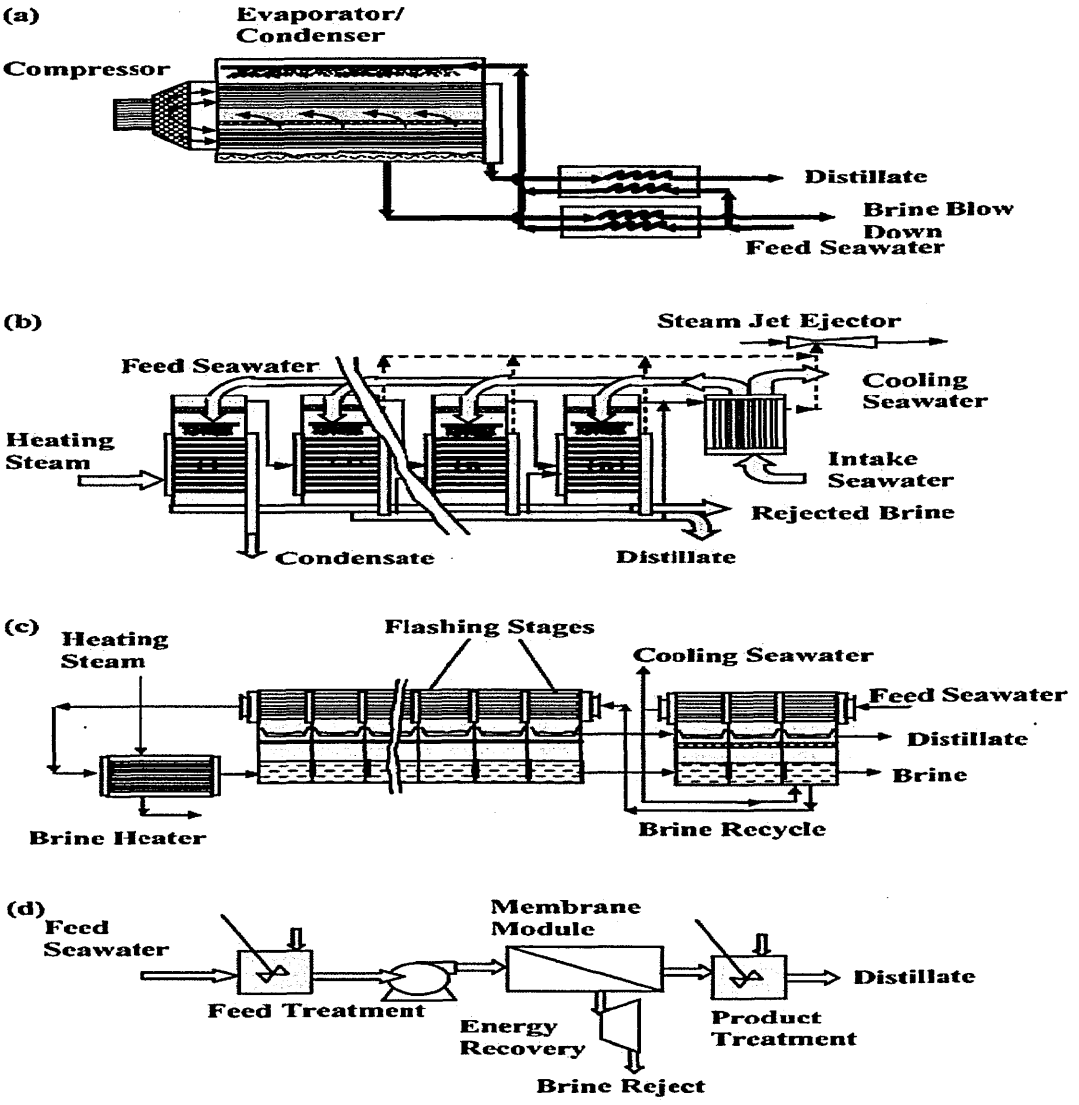


Figure 1.6: Schematics of conventional processes. (a) MVC, (b) MEE, (c) MSF, and (d) RO (13).

1.4.3 Other Methods

Tsiourtis (10) presented the other methods as: freezing, membrane distillation, solar distillation and wind-driven processes. These processes have not achieved the same level of commercial success, but they may prove valuable under special circumstances or with further technological advances or improvements.

1.4.4 Future Desalination Technologies

The future desalination technologies are electro-deionization (EDI) which uses ion exchange with electro-dialysis, Membrane Distillation (MD) which uses the temperature difference between the sides of the membrane, and Freeze Separation (FS) which uses separation of pure water ice from the brine of seawater (14).

1.5 Desalination Requirements

Desalination requires the following (10):

- Saline or brackish water must be of acceptable quality of contamination.
- Chemicals additives for pre-treatment and post-treatment and must be safe for the environment.
- Chemicals for anti-scaling, anti-corrosion and oxygen scavenging.
- Thermal and electric energies for thermal water distillation and RO respectively.
- A suitable land for the construction of distillation plant.

1.6 Desalination Output

The desalination process has the following emissions

- Brine which has a high salt concentration.
- Desalinated water.
- The residual products of chemical additives.
- Scale and sludge.

1.7 Selection of Desalination Process

The selection of the desalination process should be based on the following parameters (10):

1. ***Raw water qualities:*** Brackish water and water from beach wells have lower concentrations of salt than seawater which will affect the choice of desalination process. RO is more suited to the lower salt concentrations whereas thermal processes may be more economic for seawater which requires additional pre-treatment such as antifoaming agents.

2. ***Product water qualities:*** Very pure water with almost zero salt content can use the distillation method or double pass membrane method is required.

3. ***Source of energy:*** If thermal energy is available from a power station, then the thermal process is preferable. Other forms of energy are electrical, solar, wind and nuclear.

4. ***Location of the plant:*** The power supply source and the type of power available in the location are very important parameters.

5. *Size of the plant:* Small plants can use RO whereas large desalination plants may be combined with power station generation and choose a thermal process.

6. *Environmental requirements:* Physical and chemical impacts of all gas emissions, noise levels, concentrations of rejected seawater, and pollutants must be considered.

Chapter 2

Literature Survey of Scale Formation

Chapter 2

Literature Survey

2.1 Introduction

In order to increase the distillate production and the thermal efficiency of desalination plant higher top brine temperatures (TBT) are used which makes scale formation more acute. To control and minimise scale formation on the heat transfer surfaces a range of chemical antiscalants have been developed. The chemical treatments and antiscalants used in MSF plant are described in more detail in Appendix C.

This chapter presents an overview of the types of scale, the mechanisms of inhibition and scale control in the desalination process. The study will also consider operational aspects with reference to the on-line mechanical cleaning of heat exchanger tubes by sponge rubber balls (Appendix D).

Scale is a mineral deposit of inorganic salts whose solubility is temperature dependent. Hard scale is formed when the temperature increases. It is deposited onto tubes of the distiller, and thus it causes poor plant performance.

2.2 Scaling

Scale is formed by deposition of seawater salts that go beyond their solubility limits. This formation of scale is inhibited mainly by the addition of antiscalants that have the characteristic to increase the threshold limits of saturated seawater salt solutions and have the ability to distort the scale crystalline lattice so as not to adhere to heat transfer surfaces (15).

Scale deposition on heat transfer tubes is an inevitable consequence in MSF evaporators of seawater desalination plant. The scales consist mainly of calcium carbonate and magnesium hydroxide and occasionally of calcium

sulphate. The scaling induces insulation in heat exchanger tubes resulting in lowering of heat transfer efficiency and in consequence, significant losses in production and high-energy consumption (16).

Scale is likely to precipitate inside the MSF condenser tubes, thereby lowering heat exchange, promoting corrosion and finally blocking or wearing tubes down. Furthermore, high operating temperatures increase the risk of scaling in MSF plants and the presence of nuclei for crystallization is also critical to initiate scale deposition (17).

A high TBT is desirable in a multistage flash distiller, despite scale formation, as it allows for more stages to be implemented, thereby increasing the area available for heat transfer and improving the Gain Output Ratio (GOR) of the plant. In a typical desalination plant, heat transfer areas are extended by up to 40% to allow for scaling and bio-fouling, which counteract and reduce heat exchange across tube walls (18).

For example, the last stage before blow-down is prone to scaling, although temperatures are relatively low, the brine salinity is high and should be limited to 70,000 ppm to avoid over saturation (19).

2.2.1 Alkaline Scale Formation

Alkaline scale such as CaC_{03} and Mg(OH)2 become less soluble with increasing temperatures as bicarbonate thermally decomposes and thereby gives way to the scale forming species. Calcium sulphate is also inversely soluble but precipitation starts at higher temperatures (>115 °C) than observed for alkaline scales and forms very hard scales in MSF plants that are difficult to remove (20). The two methods of minimizing alkaline scale formation in MSF distillers are:

1. Acid dosing type (pH control).
2. Antiscalant dosing chemical to the make-up water seawater (21).

Areas of scale in distillers are:

1. Brine Heater.
2. Demister meshes.
3. Water tube sheets.
4. Flash chamber orifices (21).

The non-alkaline scale formation (mainly $CaSO_4$) in MSF distillers is controlled by maintaining the TBT below 122°C in order to limit its formation. The suggested mechanisms by Shams El Din et al. (22) are:

$$2HCO^- \rightarrow CO_3^{2-} + CO_2 + H_2O \qquad 2.1$$

As temperature is increased further carbonate ions hydrolyze by the following reaction:

$$CO_3^{2-} + H_2O \rightarrow 2OH^- + CO_2 \qquad 2.2$$

A uni-molecular decomposition of bicarbonate along with its neutralization is given by the following reactions:

$$HCO_3^- \rightarrow CO_2 + OH^- \qquad 2.3$$

$$HCO_3^- + OH^- \rightarrow CO_3^{2-} + H_2O \qquad 2.4$$

It was also suggested that there would be subsequent precipitation of $CaCO_3$ and $Mg(OH)_2$. The appearance of either scale at lower temperature stages of the MSF distiller by one mechanism or the other could be attributed to circulation with the brine recycle (23) , (21).

Glater (24) and Harris et al. (25) in their papers on magnesium scales suggested the following order of reactions:

$$HCO^- \rightarrow OH^- + CO_2 \qquad 2.5$$

$$Mg^{2+} + 2OH^{-} \rightarrow Mg(OH)_2 \qquad 2.6$$

$$HCO_3^{-} + OH^{-} \rightarrow CO_3^{2-} + H_2O \qquad 2.7$$

$$Ca^{2+} + CO_3^{2-} \rightarrow CaCO_3 \qquad 2.8$$

Mubarak (23) states "It is clear that the ambiguity revolves around the underlying kinetics and since the consequences of adopting either concept would, no doubt, have a great impact on the operational as well as maintenance philosophy of desalination plants, the need for a kinetic study that will hopefully contribute to the understanding of such a complex system is therefore readily apparent".

Scale formation, which impairs heat transfer and reduces flow velocities inside the condenser tubes of desalination plants, is a major operational problem. Alkaline scales result from the interaction between the decomposition and hydrolysis products of the bicarbonate ion of seawater with calcium and magnesium ions to produce $CaCO_3$ and $Mg(OH)_2$ (26).

2.3 Scale Formation and Prevention

Scale formation is, together with corrosion, the most critical phenomenon of the desalination process. It represents the main cause of the decrease in performance and availability of MSF plants and, being the most important reason for the upper limit to the brine temperature, is also the major bottleneck for plant performance improvement.

2.3.1 Alkaline Scale

In thermal seawater desalination plants, particularly in MSF process, the phenomenon of scaling takes place normally through the formation of "alkaline scale" (calcium carbonate and at higher temperatures magnesium hydroxide) shown in figure 2.1. Their solubility is inversely proportional to the temperature

and will precipitate as they exceed their maximum solubility in the region of highest temperature. This leads to scale formation in crystalline form on the metal surface, which acts as a resistance to heat transfer.

Figure 2.1: Sample of alkaline scale.

The rate of formation of calcium carbonate and magnesium hydroxide in seawater depends on temperature, pH, concentration of bicarbonate ions, rate of CO_2 release, concentration of Ca^{2+} and Mg^{2+} ions and total dissolved solids.

Calcium carbonate is the major scale that is formed up to a seawater temperature of 90°C, while magnesium hydroxide is mainly formed above temperatures of 95-100°C. This is due to the increase in formation of hydroxyl ions with increasing temperature. It is to be noted that calcium carbonate and magnesium hydroxide have an inverse solubility; with an increase in temperature their solubility decreases.

2.3.2 Hard Scale

The other form of scale formation, non-alkaline, that can be encountered in MSF units is due to calcium sulphate precipitation, usually known as 'hard scale'. It is to be remarked that the precipitation of calcium sulphate in MSF plants has very severe consequences, as this is a deposition that is very difficult to remove in tube exchangers. Its occurrence is due to incorrect plant operation leading to excessive brine concentration and temperature.

Calcium sulphate exists in three different forms: anhydrite ($CaSO4$), hemi-hydrate ($CaSO4.1/2H20$) and di-hydrate ($CaSO4.2H20$). All these are more soluble in seawater than calcium carbonate and magnesium hydroxide, and the solubility of both anhydrite and hemi-hydrate decreases as temperature increases. Precipitation of $CaSO4$ and its hydrates is independent of pH, and is affected by the concentration of the other ions in the solution.

The three different crystalline forms differ markedly in solubility, each having its own respective solubility range (the least soluble being anhydrite and the most soluble di-hydrate).

The essential facts to consider in tackling the issue of hard scale prevention in seawater desalination are the following:

- Anhydrite is stable in seawater above 100°C and di-hydrates below 100°C.
- Hemi-hydrate is the meta-stable form of calcium sulphate.
- The transition from hemi-hydrate to anhydrite is very slow.

2.4 Scale Control Methods

Two kinds of alkaline scale control methods are commonly used in MSF desalination practice: organic polymers addition and acid dosing in feed seawater.

Organic polymers used in the control of scaling have properties which affect the threshold on salt precipitation, crystal growth inhibition and crystal distortion (20), as shown in Figure 2.2. In principle, the action of such additives results in maintaining the calcium and magnesium ions in solution beyond the saturation value, having at the same time an effect on the crystal habit of the scaling compound, such that the crystal either stops growing or is distorted.

(a)

(b)

Figure 2.2: Box wall, tubes and tube sheet showing scale deposition in the inlet of Brine Heater with (a) using antiscalant (b) without.

The distorted crystals are less likely to adhere to each other and to metal surfaces. Polymers based on poly-carboxylic acid type chemistry also have the additional advantage of having dispersion properties of calcium carbonate, magnesium hydroxide and silt, as they can absorb onto the particulate surfaces and impart a like charge, hence repelling neighbouring particles and keeping them suspended in solution.

The chemical additives that are currently in use have one or more of the properties mentioned above. These are polyphosphates, phosphonates, polyacrilates and polymaleates.

Polyphosphates are rarely used in MSF plants, as they tend to form additional scale in the form of thick muddy deposit as calcium phosphate, which can strongly adhere to the heat transfer surfaces and is difficult to remove. The other types of additives are used on their own or in combination with a mechanical cleaning method (sponge ball cleaning), in order to remove the soft scale deposited on the heat transfer surface.

The retardation of scale formation through addition of antiscalants is usually coupled with sponge ball cycling. The diameter of the ball is chosen to be slightly larger than that of the tubes, and the balls are forced through by the pressure of the running water. As they travel through the tubes, the balls wipe out any loose scale or deposit and prevent the formation of a continuous, solid layer.

The use of acid dosing to control alkaline scale formation is based on the principle of reducing the concentration of bicarbonate ions in feed seawater down to a residual M alkalinity of about 10-20% of seawater initial alkalinity (this residual alkalinity is a safety margin to protect the plant against risks of acid corrosion). The reaction of acid ions with bicarbonate ions present in seawater causes the formation of carbon dioxide, as follows:

$$H^+ + (HCO_3)^- \rightarrow 2H_2O + 2CO_2 \qquad 2.9$$

In such a way, the pH of the re-circulating brine is maintained at a value in the range 7.2 to 7.8, depending on the CO_2 removal efficiency of the de-carbonation and de-aeration system, and the precipitation of calcium carbonate and magnesium hydroxide can occur only to the small amount of the residual alkalinity.

2.5 Scale Control of Brine Heater and Heat Recovery Section

The operation and maintenance manual for the Doha West distillation plants (35) provides guidance for the control of scale. The plant is designed to perform scale control by dosing anti-scale chemicals to the make-up seawater.

The chemical acts on the precipitation of alkaline scale so that rate of the precipitation is retarded considerably. The scale precipitates gradually with plant operating hours. The precipitation on the inside surface of heat exchanger tubes

will cause reduced performance of the heat exchanger and a decrease in thermal efficiency of the plant.

In order to prevent the formation of non-alkaline scale the brine concentration should be controlled not to exceed the solubility limits of them. Anhydride calcium sulphate scale, however, may precipitate in a tube due to its long retention time in the super-saturated condition. This may be caused by a tube which is accidentally blocked by debris or foreign material.

An on-load tube cleaning system is another means to prevent tubes from excessive precipitation of the alkaline scale. Sponge balls cleaning is also effective to control the precipitation of the alkaline scale in the tubes but is not effective for non-alkaline scale.

2.6 Scaling on Heat Rejection Section

In principle the tubes of the heat rejection section of the evaporator are free from scaling, but there is a tendency for precipitation of the alkaline scale because seawater occasionally becomes supersaturated due to scale components. As a preventative measure when the plant is shutdown, seawater in the tubes must be drained out completely. Furthermore, if the plant is out of operation for long term period, the inside of the tubes should be cleaned by flushing with fresh water.

2.7 Kuwaiti Experience in Scale Control

Al-Zubaidi described the Kuwaiti experience in scale control as following: “The Kuwaiti operators were concerned about scale formation inside the tubes of the distillation plant, especially since the recommended top brine temperature was 93°C and the chemical treatment suggested was Hagevap (PD8). Kuwait experience with scale up to 1953 was confined to scale formation on the outside surface of the tubes of submerged-type evaporators, where a brine temperature of about 82 - 88 °C is used in order to operate efficiently” (27).

“There were two vertical plates in each flash chamber to cascade the vapour before it reached the heat transfer area to condense the distillate. As well as this the vapour velocity was high especially in the fourth stage which was under a higher vacuum. The coefficient of the heat transfer specified was doubted due to the fact that scale was expected to precipitate inside the tubes” (27).

“Westinghouse agreed to introduce most of the recommended alterations into their design and to build a prototype consisting of two stages, which would be full size in cross section, and to study the following:

a) Brine control between stages.
b) Purity of distillate.
c) Scale formation tests.

The pilot plant tests for the brine control between stages indicated that orifices of 11.4 cm (4.5 in) diameter between the flash stages and at the entry of the top flash chamber was the optimum size to keep the balance of brine levels in all stages at full load and reduced load” (27).

“Trials were carried out with reduced loads and at a top brine temperature of 123 °C with a brine velocity inside the tubes equal to that for full load. Two series of tests were carried out, each of 100 hours duration. The first was with dose rate of 4 ppm and the second with 2.3 ppm of Hagevap compound. The conclusion was that Hagevap compound was suitable as an antiscalant additive but that it would not stop sludge formation. It was decided that the sludge could be cleaned using an acid treatment (27) “.

2.8 Developments in Chemical Treatment in Kuwait

Also, Al-Zubaidi explained the developments in chemical treatment in Kuwait as: “ Until 1953 all distillation plants were designed to operate using polyphosphate chemical dosing only. The utilization of this chemical, limits the

top brine temperature to about 90 °C, as above this temperature hydrolysis of the polyphosphate occurs leading to the formation of calcium ortho-phosphate sludge which readily adheres to the heating surfaces of the condensers and reduces the thermal efficiency of the unit" (27).

"When MSF distillation was first introduced in Kuwait in 1953, as shown in Figure 2.3, the chemical originally recommended by plant manufacturers was a proprietary mixture Hagevap or "PD8". This chemical proved to be successful and was used for many years starting in 1955 at a dosage rate of 4 ppm" (27).

Figure 2.3: The first power station in Kuwait (Shuwaikh Station) in 1953 (9).

"It was realised that if a chemical was used which would inhibit scale formation and at the same time allow operation above a top brine temperature of 90°C, higher plant output and greater efficiency would result. During investigations, MEW laboratories developed a mixture of chemicals which proved to be as good as the proprietary chemical (27) ".

"By purchasing in bulk quantities the constituent chemicals and mixing them, the end product proved to be cheaper than the original compound. The mixture which contains 49% sodium tri-polyphosphate, 49% lignin sulphonate and 2% anti-foam agent was used at a dosage rate of 4.5 ppm. An accurate,

continuous, uniform and homogenous dosage of this mixture was found to be essential to control scale formation (27)

In addition, Al-Zubaidi (27) stated in 1987 " Over the last sixteen years many polymer additives designed to prevent scale formation were operationally tested in order to determine their performance as far as scale inhibition is concerned". Some of the additives tested included Darex 40, Calnox 214 and Vaptreat H. These additives proved effective up to a temperature of 93°C, however, their unit price made them uneconomical. Another polymer additive, Cyanamer P-35 was also tested, but this did not prove to be as effective as the polyphosphate additive" (27).

All the trials were carried out on the Shuwaikh F plants. The next trial to be carried out on this plant was for the evaluation of a new antiscalant additive called Belgard EVN. The trial confirmed Belgard EVN to be superior in performance compared with polyphosphate- based additives. The further attraction of Belgard EVN was that it enabled operation at a top brine temperature of 105°C. In order to carry out an accurate performance comparison between polyphosphate and Belgard EVN, it was necessary to monitor the performance of polyphosphate for a reasonably long time using carefully calibrated instrumentation (27).

After having obtained reliable long-term low temperature polyphosphate data, a trial of Belgard EVN was initiated, using a dose rate of 2.0 ppm. The short term trial did prove that Belgard EVN was quite effective as a scale control additive at a dosage rate as low as 2.0 ppm. Another trial was carried out on Belgard EVN which demonstrated the capability of the additive to reduce scale deposition up to 105°C, thus enabling more distillate output (27).

It had been established that as far as scale prevention is concerned, Belgard EVN was superior to polyphosphate. However, due to the much higher unit price of Belgard EVN, a trial was carried out using two large units 6 MIGPD running simultaneously, with one unit utilizing polyphosphate and the other using Belgard EVN. The trial which lasted for a whole year was devised not only

to monitor the performance of both additives over such a long time period using high product capacity units but also to monitor the steam consumption of both units and other operational factors (27).

Chapter 3

Multi-Stage Flash Technology

Multi-Stage Flash Technology

3.1 Introduction

Most of the seawater desalination plants are of the MSF type. The basic principle of the MSF distillation process is to evaporate seawater and condense the vapour. This chapter comprises of a simplified explanation of the main sections of MSF desalination technology and also, refers to the historical review of MSF desalination technology.

3.2 MSF Desalination Technology

The MSF process evaporates seawater which it is fed into the first evaporator (flash chamber) at a lower pressure than saturation pressure. The vapour condenses and cools down to a saturation temperature equivalent to the flash chamber pressure.

The flashing gives rise to the production of a certain amount of vapour, generated until equilibrium conditions are achieved again. The intensity, duration, yield, etc. of the flashing process strongly depend on the physicochemical properties of the fluids, on the temperature and particularly on the pressure drop which causes the phenomenon itself.

Figure 3.1 shows the process on a T-s diagram. As the brine enters the stage at 1 it expands through a throttling process, reducing in pressure and temperature to 2. Some of the flashed steam will be condensed and the rest will be carried over to the next stage at 3 to continue the process.

T

Figure 3.1: T-s diagram for the flashing process

Figures 3.2 and 3.3 show views of the DWPS desalination plant and Figure 3.4 shows a general schematic of the plant.

Figure 3.2: Aerial view of the DWPS MSF Desalination Plant.

Water Box

Water Box from outlet o f stage 12 to stage 13

Distillate Header

eat Rejection outlet for stage

Figure 3.3: View of the DWPS MSF Plant.

Dearator

Distillate Line

Figure 3.4: Schematic of a typical multistage flash desalination plant (11).

Figure 3.5 shows a schematic flow diagram of the MSF desalination plant. The MSF plant consists of three main sections: heat input (brine heater), heat recovery, and heat rejection. The heat rejection and heat recovery sections consist of a number of flash chambers (stages) connected to one another.

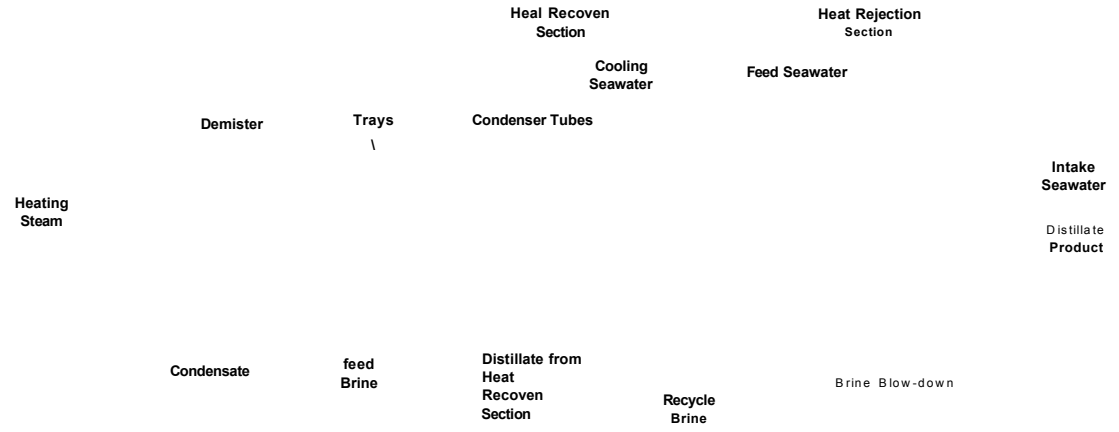


Figure 3.5: Schematic diagram of the MSF desalination plant (28).

K. A. AL-Shayji (29) explained the above process as: “Seawater enters through the heat rejection section. This section uses the heat released during condensation to preheat the feed and to reject energy into the supplementary cooling water. The re-circulating brine, which is formed by mixing part of the feed seawater (make-up) and a large mass of brine from the last stage, is circulated through heat-recovery tubes”.

In the heat recovery section, the brine is heated as it passes through the tubes from one stage to another by exchanging the thermal energy from the flashing vapour in each stage. Thus the heat released by the condensation of vapour is used to heat the re-circulating brine. Passing through the last stage, the water enters the brine heater, where its temperature is raised to a certain temperature which is equal to the saturation temperature (TBT) for the system’s pressure (29).

After heating the saturated brine to the TBT in the brine heater by the saturated or supersaturated steam coming from the boiler, the saturated brine enters the first stage of the heat-recovery section through an orifice or weir. As the brine runs into the first stage, it will become superheated and flash-off to give pure vapour as a result of the pressure reduction. The vapour then passes through the demisters, where the salt carried with the vapour is removed, condenses on the cooling tubes, and is collected as distillate in the distillate tray (29).

Figure 3.6 shows a schematic cross-section of a single stage which is called a “flashing chamber”. The previous process is then repeated as brine and distillate enter the next stage at a lower pressure. The distillate is finally collected in a distillate tray.

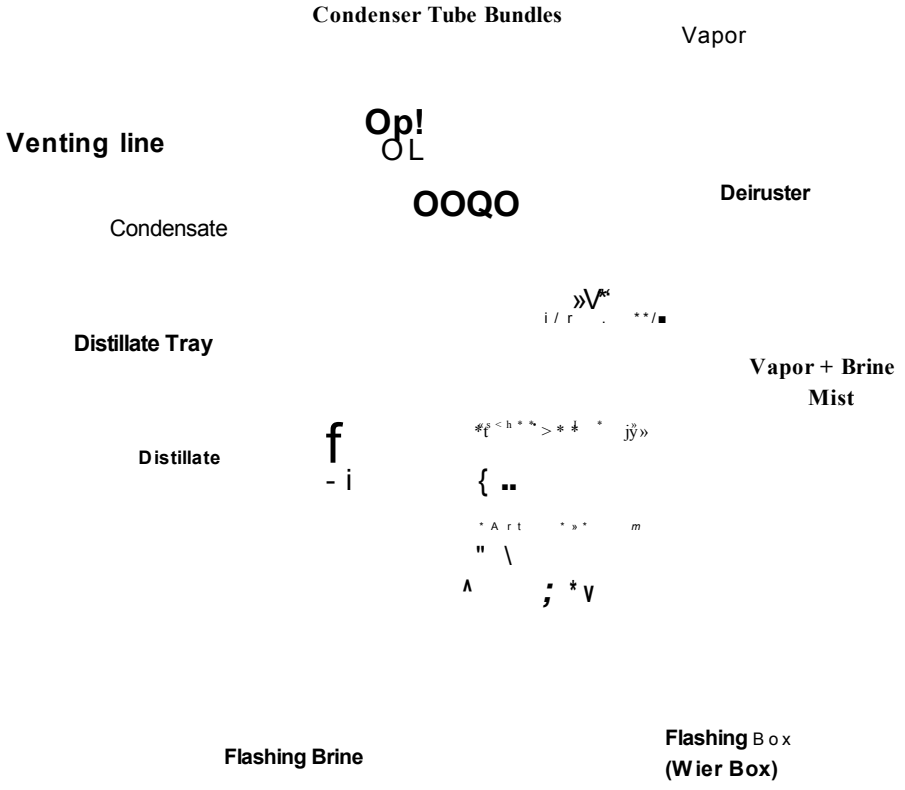


Figure 3.6 A single stage in the MSF desalination plant (29).

Each stage consists of a brine orifice, condenser tubes, pre-heater tubes, water boxes, distillate tray, a venting line, a demister and partition walls. The functions of the mentioned parts are:

- The demister reduces the amount of captured brine droplets in the flashed-off vapour, in order to prevent scale formation on the outer surface of the condenser tubes and the walls of the flashing chamber. Excessive entrainment of brine in the flashing vapour would also result in high product salinity.

- The brine orifice controls the brine flow rate across the stages. This in turn affects the process of bubble formation, growth and release rate. The height of the brine orifice is adjusted to prevent leakage of the flashed-off vapour across the stages (30).

- The heat transfer area of the condenser tube controls the temperature of the brine recirculation which enters the brine heater. Also, use of excessive heat transfer area may result in a lower velocity of the brine recirculation inside the condenser tubes, which may enhance the fouling process. The converse, the use of a smaller heat transfer area, would result in low energy recovery and high brine velocity (19).

- The intake of the venting line is located at the coolest point within the flashing chamber. The venting line provides an escape of the non-condensable gases from the flashing chamber and prevents their accumulation (19).

Finally, part of the brine from the last stage is then recycled to the heat recovery tubes after adding make-up seawater to it. The brine is heated as it passes through the tubes from one stage to another by exchanging the thermal energy from the flashing vapour in each stage, and the cycle is repeated (29).

3.3 Heat Transfer Mechanism of MSF

The heat exchange through the condenser tubes of an MSF evaporator, from the vapour phase external to the tubes to the brine flowing inside the tubes, involves various forms of heat transfer, which are briefly discussed below.

a) Conductive heat transfer through the tube wall, where the transfer coefficient is given by the general conductivity law (proportional to the thermal conductivity of tube metal and inversely proportional to tube wall thickness).

b) Fouling resistance is correlated to the scale deposition and bio-fouling that decreases the heat exchange rate of the condenser with time. It is strongly dependent on the type of scale inhibition method used, and also on the brine

temperature and velocity in tubes. A fouling resistance value appropriately selected assures that the performance of the exchanger will be at the design value after a specified period of time.

The inverse of the sum of the thermal resistances represents the overall heat transfer coefficient for the case in subject.

3.4 The Development of MSF

The modern version of multiple-stage flash distillation is comparatively new, having been invented in the late 1950s, although before that time the principle of flash distillation had been known. The earliest application of a flash process for the production of fresh water used arrangements resembling submerged-tube multiple-effect plants. Vapour, produced by "flashing" in one of the chambers, was piped to a second chamber where it was condensed on the outside of tubes through which cooling water flowed. Thus for each stage there are two separate chambers: a flash chamber for producing vapour and a condensing chamber where the vapour was liquefied. An example of this early type of flash installation is shown diagrammatically in Figure 3.7.

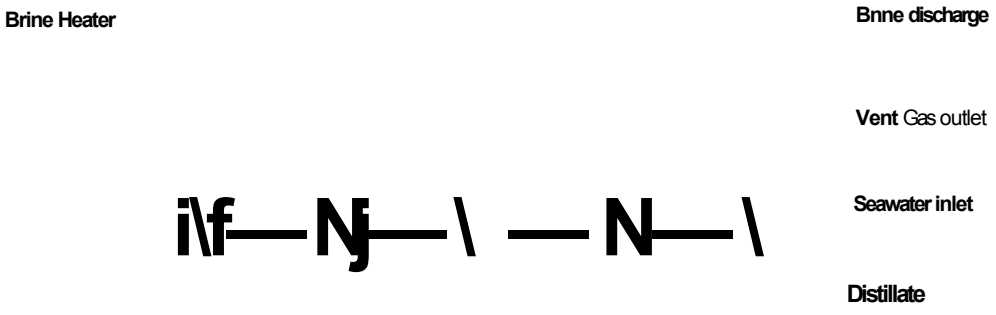


Figure 3.7: Schematic arrangement of the five-stage flash plant.

A refinement of this same type of design is encountered in a 1953 Kuwaiti installation, figure 3.8, supplied by Westinghouse, where most of the external piping is eliminated and several stages are enclosed in a single casing. The

flash distillation plant, having a total production capacity of 2.5 Million Imperial Gallons per Day (MIGPD), consists of four identical units, each having four stages. The vertical towers of the four individual units are single casings that contain the four stages as layers, one above the other, with the higher temperature stages at the top.

Figure 3.8 The first distiller at the Shuwaikh Station 1953 (9).

The first true example of a modern MSF arrangement can be found in the installation supplied by Weir Westgarth (UK) at Shuwaikh, Kuwait, in 1960. The plant was formed by two units, each having a production of about 1 MIGPD (4,500 m^3/day), with a performance ratio of 5.7 and a 19 stage evaporator (Appendix B).

3.5 MSF Process Scheme

The two main alternative process schemes that are used in MSF practice are brine recirculation and once through. Both are discussed in the next sections:

A. Brine recirculation scheme

The MSF distillation system with brine recirculation can be subdivided in the following sections:

- Heat Input Section (Brine Heater)
- Heat Recovery Section
- Heat Rejection Section

The latter two sections are usually combined in the same vessel, called the evaporator, and consist of a number of flash chambers in series; the brine (concentrated sea water) flashes at the bottom of each stage and the vapour released condenses on the outer surface of the exchanger tubes installed in the upper part of each stage.

The process on which this kind of plant operates is based on the recycle of brine in the Recovery Section, where the latent heat of condensing vapour is recovered by increasing the temperature of the brine re-circulating within the condensers tubes; in this way it is possible to obtain a high thermal efficiency of the process.

The heat input to the system is supplied by incoming steam (at low pressure) to the brine heater, a tube-bundle exchanger in which the temperature of the brine flowing inside the tube is raised by thermal exchange with condensing steam. The heated brine then passes through suitable orifices to the first evaporator chamber (Heat Recovery Section).

The vapour released flows through the demisters (where any droplet of entrained seawater is separated) and, transferring its latent heat through the stage condenser to the re-circulating brine, generates a stream of distilled water, which falls in a suitable tray below the tube bundle and flows to the next stage.

This process causes a decrease in temperature of flashing brine and a corresponding increase in temperature of re-circulating brine (flowing inside condenser tubes).

The above process is repeated (at decreasing temperatures and pressures) for all the stages of the Heat Recovery Section, by transferring un-flashed brine to the lower temperature stages.

The brine then passes to flash in the stages of the Heat Reject Section, with a distillation scheme similar to what has been described above, with the only difference that in this case the cooling medium for condensation is fresh sea water rather than re-circulating brine. In this way, the heat entered into the system in the Brine Heater is rejected to ambient at low temperature.

The salt concentration of re-circulating brine is maintained at a value of about 65,000-70,000 ppm by a continuous blow-down of concentrated brine and a congruent feed of make-up seawater, which is de-aerated and treated with antiscalant chemicals prior to entering the evaporator. Make-up is introduced into the flash side of the last reject section where it mixes with the major stream of concentrated brine that is not discharged as blow-down.

After mixing, brine in the last stage is extracted by the brine recirculation pump and delivered to the condenser tubes of the last recovery stage, where it acts as the cooling medium for the vapour generated by the flashing brine in the same stage. Re-circulating brine then passes through the condenser tubes of the next stages, in a counter-current manner, rising progressively its temperature by absorbing the latent heat of condensing vapour, as previously described.

This establishes a temperature profile increasing from the reject section, through the recovery stages, to the brine heater. The arrangement for brine transfer across the adjacent chamber is such that no vapour may leak from each chamber, which allows establishing a pressure profile through the chambers corresponding to the vapour saturation temperature.

In counter-current flow inside the exchanger tubes, re-circulating brine is finally transferred from the first stage condenser to the brine heater, where its temperature has to be raised only a few degrees to allow continuation of the thermodynamic cycle.

The temperature of brine leaving the brine heater is in the range 90°C - 120°C, while its temperature in the last stage is typically 8-1 OK above the seawater temperature. Corresponding vapour saturation temperature in the flash chambers varies from 1.9 bar (abs) to 50-100 mbar (abs), which means that most evaporator chambers operate under vacuum. In order to maintain the vacuum condition, air leakage and non-condensable gases released in the evaporator from feed seawater (essentially air and CO_2 formed by carbonates breakdown) have to be continuously purged to atmosphere.

The main pumps necessary for the process are the following:

- Condensate extraction pump, delivering to the condensate circuit the steam condensed in the brine heater;
- Brine recirculation pump, extracting brine from the last flash chamber (after mixing with feed seawater) and delivering it to the condensers of the recovery section;
- Brine blow-down pump, delivering to discharge a part of the brine in the last flash chamber;
- Distillate extraction pump, extracting distillate from a sump in the last stage and delivering it to storage.

B. Once through scheme

The main difference between the once through and the brine recirculation scheme is that, there is no recirculation of any portion of flashed brine as shown in Figure 3.9. Also in this case, the seawater entering the last stage condenser is heated by passing in counter-current to the stages at higher temperature where recovery of the latent heat occurs but, unlike that in the brine recirculation scheme, it is transferred through all the stage condensers up to the brine heater.

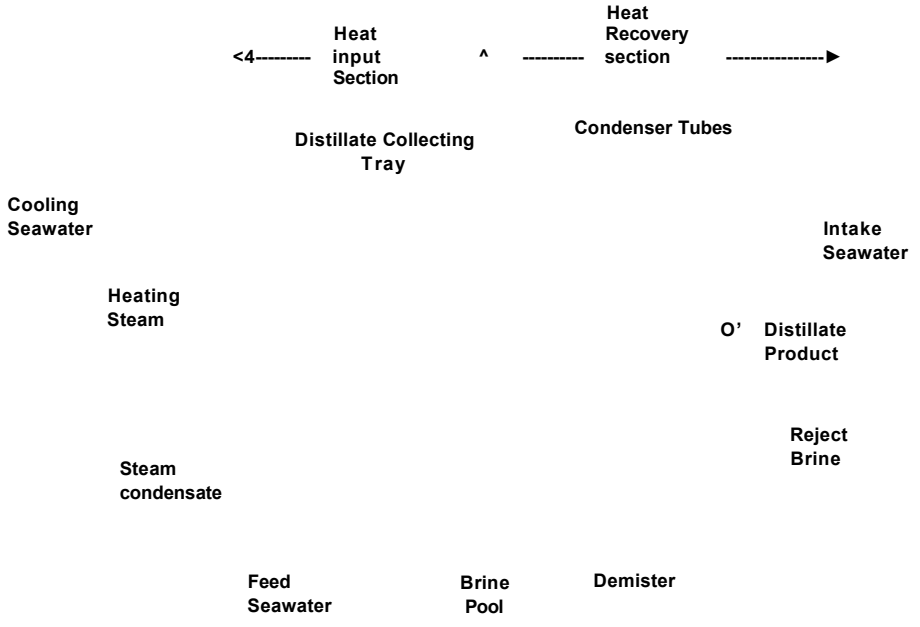


Figure 3.9: Once-through MSF desalination plant (28).

There is therefore in this case no distinction between the Heat Recovery Section and the Heat Rejection Section. After being heated by thermal exchange with condensing steam, seawater is discharged to the evaporator flash chambers, where distillate is produced similarly to what previously described. In the last chamber, all the brine is discharged as blow-down.

The main advantages of the once through scheme in comparison to brine recirculation are the reduced scale formation due to low brine concentration, a simpler operation and some cost reduction. The disadvantages are related to the large quantity of raw seawater supplied to the evaporator that implies a greater release of non-condensable gases with an associated corrosion

potential, as well as an increase in size and consumption of the vacuum system.

The once through system also presents reduced operation flexibility, as in this case it is not possible to control independently the top temperature, the bottom temperature and the distillate production. Probably due to the above constraints, the once through selection is generally limited to small plants, while most of MSF units installed are of the brine recirculation type.

3.6 Materials of Construction of MSF Plants

- **Evaporators**

Yousef Al-Wazzan (31) stated that: “The Kuwaiti specifications require minimum corrosion allowances in the flash chamber walls of 12.7 mm for the first and last flash chambers, and 9.3 mm in other chambers, and a thickness of at least 38.1 mm”, as it is shown in Figure 3.10 and 3.11 (31).

Flash Chamber

Figure 3.10: View of Flash Chambers of the DWPS MSF Plant.

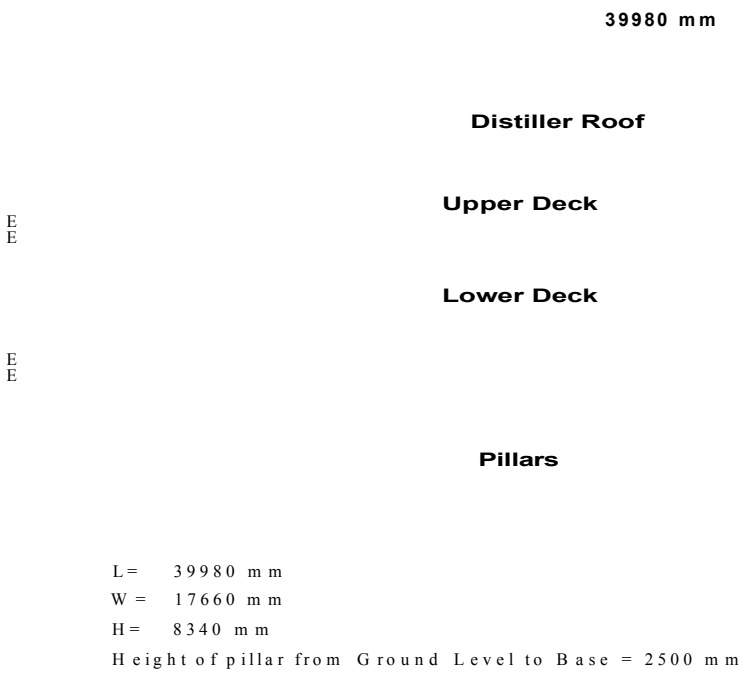


Figure 3.11: Dimensions of a MSF Distiller Unit.

- **Condensers**

Condenser tubes for the brine heater, heat rejection section, and the first and second stages of the heat recovery section are made of Cu-Ni (66% copper, 30% nickel, 2% iron and 2% manganese). Aluminium brass (76% copper, 22% zinc and 2% aluminium) tubing is used for all remaining stages (31).

- **Water boxes**

Water boxes are usually made of carbon steel clad with Cu/Ni or other corrosion-resistance metals (31).

- **Demisters**

Wire mesh made of 316 stainless steel demisters are usually used in all stages. The knitted wire mesh demisters consist of a bed, usually 10-15 cm deep of fine diameter wire interlocked by a knitting operation to form a wire mesh pad with a high, free volume usually 97-99% (31).

Product water troughs

Product water troughs are difficult to repair and to replace; consequently prepared materials should be chosen. The product troughs suffer corrosive attack from turbulent distillate water and from flashing brine (lower surface). If the vent cooling section is near the trough, non-condensable gases may also contribute to corrosion. Among the materials used for product water troughs are solid types of 316 stainless steel, 90/10Cu/Ni or carbon steel with a liberal corrosion allowance (31).

Heat exchangers designed with a corrosion allowance are typically designed to TEMA Class R, C or B according to ASME Codes. The corrosion allowance specified is 1/16" (1.5875 mm) for classes C and B and 1/18" (1.4097 mm) for class R.

The type W u-tube series includes a corrosion allowance of approximately 1/16" on all carbon steel components of the shell only. Although this corrosion allowance is not included in the ASME code calculations, it has been added to the design in order to provide a shell with a longer service life.

For stainless steel usually no corrosion allowance is specified and the corrosion allowance is being calculated based on the corrosion rate that is a function of operating conditions and type of fluid.

The Brine Heater Corrosion Allowances are: Shell side: 1.6mm (carbon steel SM41B), Tube side: 70/30 Copper Nickel C7150T and Water Box: Carbon steel 90/10 Cu-Ni Lining or 90/10 Cu-Ni clad (32).

- **Pumps**

It is recommended to use 316 stainless steel impellers, casing and shafts to obtain maximum service life in any type of desalination plants. For smaller pumps and for product water service bronze, Ni-resistant and cast iron impellers or casing are used (31).

3.7 Seawater Pre-treatment

The source of desalinated water in Kuwait is seawater from the Arabian Gulf that is connected to the Hurmoz strait and opened to the Indian Ocean. The salinity of seawater of Arabian Gulf is increasing because the flow of fresh water entering the Arabian Gulf from the Shatt Al-Arab in reducing, thus the salt content is rising. Figures 3.12 and 3.13 show the water intakes areas at DWPS and Table 3.1 lists a complete analysis of seawater at DWPS (26).

Figure 3.12: Aerial View of Sea Water Intake Area

Figure 3.13: View of Sea Water Intake Area.

Water may be purified by a number of desalination techniques in which the dissolved impurities are removed from water or, more correctly, pure water is removed from the impurities. Feed waters suitable for desalination are generally classified into four categories namely: sea water, brackish water, surface water, and ground water. Sea water is highly corrosive to iron pipes because of the high level of dissolved salt concentrations, especially sodium chloride which accounts for 60 to 70% of the total dissolved solids (TDS).

Parameter	Value
pH value	8.0-8.25
Conductivity, µS/cm	52-60
TDS, at 180°C, ppm	42,000-45,000
Total hardness, ppm as $CaCO_3$	7,600-8,000
Total alkalinity, ppm as HCO_3^-	110-130
p-Alkalinity, ppm as CO_3^{2-}	10-15
Chloride, Cl^-, ppm	23,100-24,100
Sulphate, SO_4^{2-}, ppm	3,100-3,500
Magnesium, Mg^{2+}, ppm	1,480-1,660
Potassium, K^+, ppm	560
Sodium, Na^+, ppm	11,200-14,000
Bromide, Br^-, ppm	64.9
Calcium, Ca^{2+}, ppm	440-560
Nitrates, NO_3^-, ppm	0.1
Nitrites, NO_2^-, ppm	0.005
Ammonia, NH_3, ppm	0.02
Sulphides, S^{2-}, ppm	0.1
Phosphates, PO_4^{3-}, ppm	0.1
Iron, Fe^{3+}, ppm	0.06
Total suspended solids, ppm	0.016
Specific density, kg/m^3	0-10
	1,028

Table 3.1: Chemical composition of Arabian Gulf water at the seawater intake of the DWPS power and desalination plant (26).

3.8 General Considerations on the MSF Market

Roberto Borsani (33) described the general considerations on the MSF market as: “The MSF technology for sea water desalination can only be compared with other technologies if sea water is considered as the feed to the process. In fact for brackish water the most conventional RO or ED process can give better performances with lower installation costs”.

The reasons why the MSF technology survived well against the emerging RO technology are mainly the following (33):

- *Easy operation:* Operation of MSF desalination plants is very easy and they have almost the same characteristics as a power plant with which they are always coupled. This means that operations and maintenance can be easily found and do not represent a major problem for the owner.

- *Very low performance degradation during the years:* It is recognized worldwide that an MSF plant properly operated and maintained has practically no performance degradation in terms of water production and energy consumption.

For all the above reasons the use of MSF Desalination is still the preferred desalination technology especially in those countries were the coupling of power generation and water generation is always necessary due to the high demand of both electric power and potable water.

Al-Mutaz (34) concluded in his comparative study of RO and MSF desalination plants in Saudi Arabia by stating “at the present low fuel cost, MSF will continue to be the optimum choice for large desalting plants... RO would be the correct selection if fuel costs are increased.”

3.9 Evaporator Design

3.9.1 Basic Design Specifications

The specification of an MSF plant normally includes the following basic design data:

a) Distillate production and quality

b) Top Brine Temperature

c) Thermal efficiency of the desalination unit

a) Small MSF units typically have distillate production rates in the range 1,000 - 10,000 m^3/day and the residual TDS will be < 30 ppm: If the water is required as boiler make-up or process water it is usually passed through ion-exchange units for demineralization.

In the case of MSF units devoted to the production of drinking water, the production specified for each 'phase' is typically in the order of 20-40 MIGPD (90,000 - 180,000 m^3/day), with unitary rates in the order of 10 MIGPD. In this case, the produced distillate is subject to a post-treatment to make it suitable for potable use, including a re-carbonation process to increase its carbonate hardness, pH adjustment and sterilization.

b) The selection of the value for *top brine temperature* plays a major role in the design of an MSF unit. As the production for a given heat input and brine recirculation rate is directly proportional to the flash range, and therefore to the TBT (the bottom temperature selection is usually limited within a narrow range by the design seawater temperature value and by the available amount of cooling seawater).

c) The *thermal efficiency of the desalination system* is usually specified by the Gain Output Ratio, which is the ratio between the mass flow rate of produced distillate and inlet steam. This parameter is usually selected by the end-user based on economic considerations, looking for the value leading to the minimum cost of produced water over the life of the plant.

3.9.2 Selection of Condenser Tube Configuration

A major construction difference of flash evaporators that merits special consideration and discussion is the orientation of condenser tubes with respect to the direction of the flashing brine flow. The 'cross tube' construction, as the name implies, has the tubes oriented in a direction perpendicular to the brine flow, whereas the 'long tube' evaporator construction has the condenser tubes installed parallel with the flow of flashing brine.

A comparison between the two types of construction leads to the following observations:

The number of tube expansions in the latter case is consequently lower. This implies a reduction in construction cost and lower power consumption, due to the reduced head of the brine recirculation pump.

a) In the long tube configuration the length of the various stages and consequently the heat exchange surface may be gradually varied according to the thermodynamic requirements, while the constructive constraints of a cross tube evaporator do not allow any variation of heat exchange surface from stage to stage.

b) The stages are narrower and longer for the long tube in comparison with the cross tube configuration, with higher brine depth but longer residence time. As a result, the flashing efficiency is not significantly different for the two cases.

c) The tube bundle is confined within each chamber for the cross tube, while in the long tube construction the tubes have to cross the inter-stage walls through suitable holes. Some vapour leakage necessarily occurs through the annular clearance between tube and hole, which marginally decreases the thermal efficiency of the system. The Doha West and Az-Zour power stations use the cross tube bundles.

3.9.3 Selection of Optimum Number of Stages

One of the first selections that the designer has to carry out when studying a new MSF plant is the number of evaporator stages. Finding the most technically suitable and cost effective solution is not a trivial problem, as it involves thermodynamic, installation and fabrication issues.

3.9.4 Design of Brine Transfer Weirs

For the efficient operation of the MSF evaporator, it is necessary to provide each chamber with a specially designed orifice that allows brine to cascade from stage to stage, assuring at the same time a hydraulic seal between two adjacent chambers, necessary to avoid vapour leakage. Brine flash occurs mainly at the outlet of such orifices (called 'brine transfer weirs' or 'brine boxes') while, depending on the geometry, some vaporization may take place inside them.

The objectives in designing the brine transfer weirs of a MSF evaporator are the following:

a) Assuring a path to the flashing brine to reduce as much as possible the non-equilibrium losses.

b) Minimizing the brine level on the bottom of the evaporator.

c) Avoiding vapour leakage between adjacent stages at all operating conditions of the evaporator, with a self-adjusting orifice design.

d) Providing high turbulence to the transferred brine in order to allow mixing of all layers, so as to assure complete vaporization in the downstream stage.

e) Simplifying construction.

3.9.5 Non-Condensable Gases (NCG) Vent System

The non-condensable gases (NCG) in an evaporator have the following origins:

a) Air leakage through the flanged connections of the evaporator;

b) Carbon dioxide released from the breakdown of bicarbonates present in the feed seawater;

c) Air dissolved in feed seawater.

As the evaporator operates under vacuum, air and carbon dioxide have to be removed continuously to maintain this condition. The effectiveness in removing gases from the evaporator is essential not only to avoid pressure increase, but also in view of plant integrity and duration, since even small quantities of oxygen and carbon dioxide present in the evaporator are the most frequent causes of corrosion attack.

The NCG are a small proportion when compared with the water vapour in the evaporator, and the designer has to pay particular care in avoiding extracting an excessive amount of vapour together with the NCG. For this reason, the condenser tube bundles are usually designed to allow the gas mixture to cool beneath the vapour saturation temperature after which most condensation has occurred, so that the partial pressure of the NCG increase and their percentage in the mixture become significant at the extraction point.

3.9.6 Demister Design

One of the critical issues in designing all types of distillers is the prevention of the carryover of saline water into the condensing section of the plant. This occurs because the evaporation process is a very turbulent one and produces a mist or fog of water drops, which tends to be carried along with the vapour due to its viscosity and motion. These drops can collect in the layer of distillate on the cooling surface of the condenser, thereby contaminating it.

The type of separators now used in most MSF evaporators consists of many layers of wire mesh, each layers being staggered relative to the next. These mats are placed horizontally and the stream of vapour rises vertically through them. As the vapour rises, the droplets collect on the mesh wires, merge into larger drops and drip from the bottom layer, the drop size being too large to be entrained by the rising vapour.

3.10 System Specifications

The MSF system used in the trials had 24 flashing stages (21 in the heat recovery and 3 in the heat rejection section). System characteristics are shown in Table 3.2, and include the heat transfer area, the number of tubes, the design fouling factor and heat transfer coefficient, temperatures, flow rates, and pressures of the major streams.

Variable	Brine heater	Heater recovery section	Heat rejection section
No. of stages	1	1 to 16	17 to 19
Logarithmic mean temperature difference, K	13.36	4.684	9.28
Tube material	90/10 Cu/Ni	1-2 (90/10 Cu/Ni) 3-16(Al-Brass)	90/10 Cu/Ni
Outer diameter of tube, mm	19	19	25
Average tube thickness, mm	1	1	1
Number of tubes / stage	1417	1510	1297
Heat transfer area / stage, m^2	426.5	599	677
Overall heat transfer coefficients:			
Clean conditions, kW/m^2K	4.519	4.612	3.728
Design conditions, kW/m^2K	2.060	2.728	2.249
Fouling factor, m^2 K/kW	0.264	0.150	0.176

Table3.2: Design parameters of the MSF system (27).

3.11 Operating Conditions of MSF Plants:

3.11.1 The Top Brine Temperature, 90-120°C

- Operating range of the antiscalant additives.
- Lower specific heat transfer area.
- Lower flow rates of intake seawater and brine circulation.
- Lower flow rates of antiscalant.
- Lower pumping power.
- Use high pressure heating steam.

3.11.2 The Vapour Velocity

- Limited to a maximum value of 4 m/s to avoid entrainment of brine droplets and reduction of the product quality.
- Higher vapour velocities result in re-entrainment of the brine droplets which accumulates in the demister pad.
- In the heat rejection section, the low stage temperature reduces the pressure and increases the specific volume. As a result, the vapour velocity is the highest among all stages.

3.11.3 Brine Tube Velocity

- An increase in the brine heater tube velocity.
- High velocity reduces fouling rates.
- High velocities increase pressure drop and pumping power.
- Tube fouling reduces the net flow area and result in higher brine velocity.
- Limited to 4 m/s with 90/10Cu/Ni and 5 m/s with 70/30 Cu/Ni, practical limits are within 1.5-2.3 m/s.
- Titanium and high steel alloys have higher limit of 20 m/s.

3.11.4 Performance Parameters of MSF Plants

- The most important variables controlling the desalinated water process are: thermal performance ratio, specific heat transfer area, specific flow rate of cooling water.
- Specific heat transfer area must be from 200 to 300 $m^2/(kg/s)$.
- Higher values for the specific heat transfer area indicate scaling in the heat recovery section and fouling in the heat rejection section.
- Increase in the specific flow rate of cooling seawater is indicated by the amount of heating steam (due to fouling).

3.12 Features and Specifications of the MSF Plants

H.T. El-Dessouky (35) summarizes the general features of the MSF plants as following:

- Stable and reliable operation by insuring adequate heat transfer area, suitable materials, and appropriate corrosion allowance.
- The MSF desalination units operate with dual-purpose power generation plants. Design of the co-generation plants allows for flexible operation during peak loads for power or water.
- Cross tube designs are simpler to manufacture and install than long tube arrangements.
- All auxiliaries that are motor driven have better operating characteristics than turbine driven units; even for the large brine circulation pumps.
- Additive treatment is superior to acid treatment, where acidic solutions may increase corrosion rates of tubing, shells, and various metallic parts.
- Appropriate system design should allow for load variations between 70-110% of the rated capacity.

- Specifications for MSF plants constructed in Kuwait are based on the following:

  - Production capacity is 6 MIGPD at an intake seawater temperature of 32°C with a minimum of 14°C.
  - Design allows for an increase in capacity to 7.2 MIGPD when special antiscalants which can operate up to a temperature of 110°C are used.
  - Optimum performance ratio is 8.
  - Top brine temperature is 90°C with polyphosphate and 110°C with polymer additives to limit scale formation.
  - Salinity ratio is 1.5 between circulating brine (67,000 ppm) and feed seawater (45,000 ppm).

Brine circulation ratio is 12.7 at 90°C; ratio is lower at higher top brine temperatures.

Brine flow rate per width of flashing chamber 810,000 kg/hr at rated design capacity. This is necessary to control the brine level and minimize the non-equilibrium losses.

Extra heat transfer area is added to provide for fouling factor, values of 0.264m2K/W, 0.150m2K/W, and 0.176m2K/W in the heat input, recovery, and rejection sections, respectively.

A minimum number of stages of 24 arranged in two tiers.

Feed water is screened with a size of 3.2 mm hole strainer.

Sodium bi-sulphite is injected in brine recycle to reduce the oxygen level to zero.

Chlorine at 2 ppm is injected in intake seawater to control marine bio-fouling.

Tube internal diameter is not less than 32 mm and tube wall thickness is not less than 1.22 mm, to prevent partial or full clogging, which promotes scaling and corrosion.

Brine velocity in the heater is not less than 2.0 m/s and not more than 2.13 and in the evaporator it is not less than 2.1 m/s and not higher than 1.8 m/s. This is to reduced residence time and the promotion of scaling, to prevent erosion and to reduce pressure drop and the subsequent increase in consumption of electric power.

Inlet steam is slightly superheated at zero bar gauge, temperature is controlled by a de-super-heater and throttling valve. Internal impingement plates and baffles are used to equally distribute the steam in the brine heater.

Brine pressure at the heater outlet must not be less than 0.69 bar gauge to prevent local boiling.

Brine and distillate flow is self-regulating throughout the plant by ensuring correct design of orifices, nozzles and weirs, which are designed to allow for 20% above full rated output.

The brine orifice allows for correct flashing with minimum pressure losses at the lower temperature stages and higher pressure losses at the higher

temperature stages. The orifice prevents splashing in the higher temperature stages and flooding in the lower temperature stages.

Brine levels must be as low as possible with a value of 0.5 m (no greater than 49 cm or no less than 55 cm)

Demisters thickness is 15 cm, securely fixed, simple to dismantle for examination and cleaning.

Demister height from stage bottom is 2.13 m to allow for large brine droplets to fall back into the brine pool.

Vapour release velocity from the brine surface is limited to 4 m/s, which is controlled by correct design of the cross section area of (length and width) of the flashing stage.

Vapour velocity in the demister is limited to 2 m/s in the first stage and 9 m/s in the last stage to limit brine re-entrainment.

Distillate purity is not more than 30 ppm from the last stage for all load conditions and seawater temperatures.

De-aerator removes all gas (carbon dioxide, oxygen, and nitrogen). The de-aerator operates in counter-current mode and contains non-reactive packing material, seawater spray nozzles, and stripping steam injection. The outlet seawater contains less than 100 ppb of 0 2 and zero carbon dioxide. Adding sodium bi-sulphite reduces further oxygen content to values below 20 ppb.

Gas venting is necessary to prevent gas build-up and reduction in the heat transfer coefficient.

Shells must withstand 2.07 bar gauge hydraulic pressure measured at the bottom of the shell (shell height is 8.3 m).

Water boxes must ensure smooth flow, avoiding turbulence, stagnation points, and be easily removable from the outside.

Two independent air ejectors for removal of air and non-condensable gases from the main heater, heat recovery, or heat rejection sections. The air ejectors must be provided with a straight tube single pass condenser, which is cooled by the distillate product from the evaporator. One quick start air ejector is used for system venting during start-up.

- All pumps are constant speed driven by Alternating Current (AC) electric motors with a minimum of 110% of the rated product capacity at fouled conditions. All pumps have stable operation with no cavitations.
- The water supply pump has a rated capacity of 12500m3/hr, with a no-flow head pressure of 4.9 bar g, a vertical spindle, and axial or mixed flow.
- The brine recirculation and distillate pumps have vertical spindles are of the axial or mixed flow types.
- The heater drain pump is horizontal type and the brine blow-down pump has a vertical spindle and is of the axial or mixed flow type.

3.13 MSF Flashing Stage

Also, H.T. El-Dessouky (35) stated that: For conventional MSF systems, the flashing stage has dimensions of 18x4x3 m in width, height, and length and it includes the following:

- A large brine pool with similar width and length of the flashing stage and a depth of 0.2-0.5 m.

- A brine transfer device between the stages is designed to seal the vapour space between the stages and to enhance turbulence and mixing of the inlet brine stream.

- The demister is formed of wire mesh layers and a supporting system. The demister function is to remove the entrained brine droplets from the flashed-off vapour. This is essential to prevent increase in the salinity of product water or scale formation on the outer surface of the condenser tubes.

- Distillate tray, where the condensed distillate product is collected and cascade through the stages. The distillate product is withdrawn from the tray of the last stage.

- Water boxes at both ends of the tube bundle to transfer the brine recycle stream between adjacent stages.

Chapter 4

Experimental Procedure

(Methodology)

Experimental procedure (Methodology)

4.1 Introduction

The Ministry of Electricity and Water (MEW) stated a policy requires that every available antiscalant chemical must be properly tested on a distillation unit, under its full control, which is of modern design, of representative capacity and equipped with adequate instrumentation (Appendix F) Only after the completion of these tests to MEW satisfaction, will the chemical additive be considered on a commercial basis. The present thesis aimed at gaining an in-depth understanding of the chemical effectiveness of the antiscalants by means of field trials to last for a minimum of three months.

4.2 Methodology

In order to monitor accurately the operation of the distillation unit, extensive plant data were collected as described below:

4.2.1 Sources of Data

Data which was displayed in the central control room was collected by the staff of MEW. This included local thermo-well temperatures, in addition to some local pressure readings and chemical drop levels. Chemical analyses were also performed by the chemical laboratories (Appendix E).

4.2.2 Data Collection

Operational data were analysed on a regular daily basis. A specific trial day starts at 07:30 am and ends at the same time the following day. MEW staff adopted collection data four times a day at 08:00, 12:00, 16:00 & 24:00 hours.

A number of parameters had to be determined which enable the evaluation of the overall performance of the distillation unit and, thus, the effectiveness of the tested antiscalant chemical in preventing and controlling scale deposits formation. All routine chemical analysis for control of the operation was performed on a regular daily basis by station chemists. The objective was to confirm that the test was performed according to the required test conditions.

The field trials were carried out on desalination units of DWPS and AZ-Zour. The instrumentation; thermocouples, pressure gauges and flow-meters were calibrated at the start of each trial and were monitored throughout the trial, as shown in figure 4.1 schematically, and detailed in table 4.1.

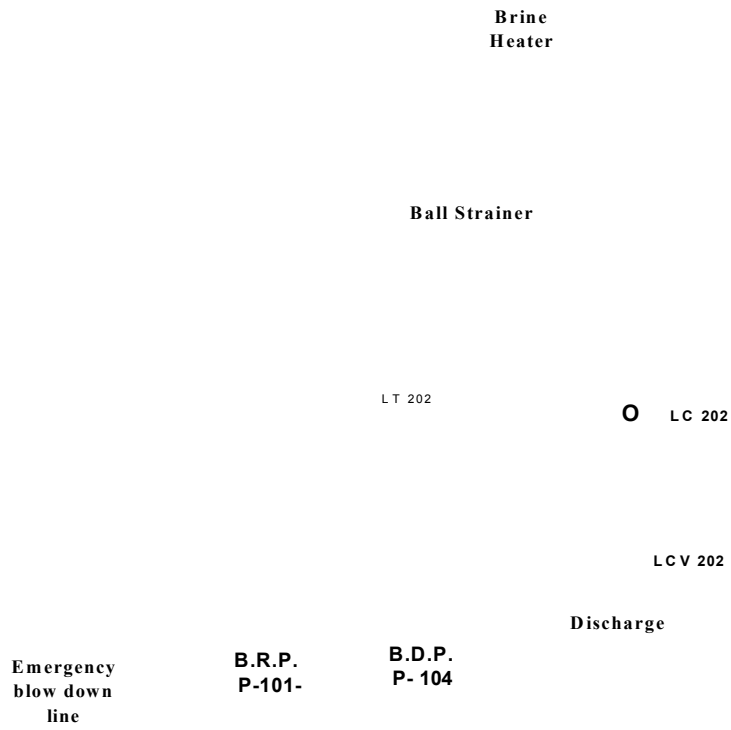


Figure 4.1: Brine Re-circulating system instrument devices (36).

Item	Description
TA311	Brine Inlet Temperature Alarm
TI 311	Brine Inlet Temperature Indicator
PT311	Brine Inlet Pressure Transmitter
PI 311	Brine Inlet Pressure Gauge
LT 202	Evaporator Last Stage Brine Level Transmitter
LC 202	Evaporator Last Stage Brine Level Controller
ZFL321	Brine Blow-down Minimum Flow Valve Limit Switch
LCV 202	Evaporator Last Stage Brine Level Control Valve
BRP 101	Brine Recirculation Pump
FV 301	Brine Recirculation Minimum Flow Valve
HCV 301A	Brine Recirculation Pump Discharge Valve A
HV 321	Emergency Blow-Down Motorized Valve

Table4.1: Description of tagged items of Figure 4.1.

Before each trial, the brine heater and heat recovery section tubes were acid cleaned. In addition, the heater tubes were also mechanically cleaned by hydro-jetting.

During the trials it was important to control the top brine and bottom brine temperatures, so that any increase in the brine heater shell pressure and steam condensate, and decrease in the heater inlet temperature, were due to fouling of the heat transfer tubes and not due to fluctuations in the operating conditions.

The effectiveness of the antiscalant additive was judged by monitoring all aspects related to the overall performance of the unit and the impact the additive might have on the operation of the unit and its efficiency.

Distillate production is tabulated on a daily basis to verify that there will be no drop in production at constant TBT and brine recirculation flow. The Gain Output Ratio, Heat Transfer Coefficient and Fouling Factor were the main focus of the plant monitoring as this gave a good indication of fouling in the distiller.

The average daily readings from the computer generated log sheets were the basis of the data evaluation. These data were used in the calculation of the

Heat Transfer Coefficient, GOR and Fouling Factor and the results were the basis of the daily judgment if uncertain trends occurred and likewise initiated measures to rectify the plant performance.

The Fouling Factor represents the resistance due to scale accumulated on the surface. If the effect of non-condensable gases is not accounted for in the condensing heat transfer coefficient, it should be included in the fouling factor.

The realistic fouling factor is difficult to obtain beforehand. The data collected from existing plants on the heat transfer coefficient, and the calculation of the clean case one are usually used to determine the fouling factor. When consultants specify liberal values for the fouling factor, more heat transfer surface than needed could be specified. When the recent sponge ball cleaning system was applied in the plant, the FF was significantly reduced.

4.3 Overall Heat Transfer Coefficients and Fouling Factors

4.3.1 Brine Heater

Heat transfer coefficients and fouling factors equations for the brine heater were calculated thermodynamically by the following equations:

$$U = \frac{W_{br}}{3600} \times 10^{3} \times C_{p} \times \ln \frac{T}{T_{c} - T_{0}} \qquad (3.1)$$

Cp determined using the average temperature $\left(\frac{T_{i} + T_{o}}{2}\right)$

$$Cp = (0.000234 \times \left(\frac{T + T_{2}}{2}\right) + 0.9245) \times 4.1868 \qquad (3.2)$$

Where:

Brine Heater Outlet Temperature, To (K)
Brine Heater Inlet Temperature, Tj (K)

Brine Heater Condensate Temperature, Tc (K)

Re-circulating Brine Flow rate, W_{br} (tonnes/h)

Specific heat of Re-circulating Brine, Cp (kJ/kg K)

Brine Heater Heat Transfer Area, A = 3544 m2

Heat Transfer Coefficient, U (kW/m2K)

Design heat Transfer Coefficient, Uc at High Temperature = 4.519 (kW/m2K)

4.3.2 Sample Calculation, Table 4.2 and 4.3 :

W_{br} = 11931 tonne/h

T0 = 107.9 °C

T, = 100.6 °C

Tc = 117.8 °C

A = 3544 m2

$$Cp = (0.000234 \times \frac{100.6 + 107.9}{2} + 0.9245) \times 4.1868$$

= 3.973kJ / kg K

$$U = \frac{11931 \times 1000 \times 3.973 \times \frac{1}{3600} Ln\left(\frac{117.8 - 100.6}{117.8 - 107.8}\right)}{3544}$$

U = 2.052 kW/m2K

Fouling Factor, FF $\frac{U \quad U,}{}$ (3.3)

where Uc= 4.519 kW/ m2K

Fouling Factor = 0.266 kW/m2K

4.3.3 Heat Transfer Coefficient and Fouling Factor, Heat Recovery Section

Heat transfer coefficients and fouling factors equations for the heat recovery section are calculated by the following equations when the operating conditions are determined from the design figures:

$$U_{Rec} = \frac{W_{br} \times 10^3 \times C_P \times (T_1 - T_{21})}{77204 \times \dfrac{[(T_{v1} - T_1) - (T_{v21} - T_{21})]}{Ln\left[\dfrac{T_{v1} - T_1}{T_{v21} - T_{21}}\right]}} \qquad (3.4)$$

Heat Transfer Area, Recovery Section = 77204 m^2

Where U_{Rec} is the overall heat transfer coefficient based on the log mean temperature difference between condensing steam and brine re-circulating.

T_1 = Re-circulating Brine Stage No.1 temperature (°C).
T_{21} = Re-circulating Brine Inlet 21 Stage temperature (°C).
T_{V1} = Vapour Temperature of Stage No.1 (°C).
T_{V21} = Vapour Temperature of Stage No.21 (°C).

C_p = Specific Heat at average temperature.

Heat Transfer Coefficient: U_c at High Temp= 4.6119 kW/m^2K

4.3.4 Sample Calculation, Table 4.2 and 4.3 :

T_1 = 367.95 K = 94.8 °C
T_{21} = 307.85 K = 34.7°C
T_{V1} = 372.00 K = 98.85 °C
T_{V21} = 313.00 K = 39.85 °C

Cp (at average temperature) = $(0.000234 \times \left(\frac{T_1 + T_{21}}{2}\right) + 0.9245) \times 4.1868$

= 3.934kJ/kg K

$$\hat{}Rec = \frac{13130.27 \times IO\,3 \times 3.934 \times (94.8-34.7) \times (1/3600)}{77204 \times ((98.85 - 94.8) - (39.85 - 34.7))}$$

$$Ln\ \frac{(98.85-94.8)'}{39.85-34.7}$$

= 2.437kJ/kg K

$$\text{Fouling Factor} = \frac{}{}_{Rec} \qquad (3.5)$$

where Uc = 4.6117 kW/m2K

Fouling Factor = 0.1935 m2 K/kW

Parameter	Equation	Design value	Unit
Flashing Range	Outlet BH Temperature - Bottom Brine Temperature	72.76	°C
Total Temperature Range	Outlet BH Temperature - Inlet Cooling SW Temperature	93.5	°C
Recirculation Ratio	Re-circulation Brine Flow / Distillate flow	8.82	-
Concentration Ratio	Re-circulating Brine Salinity (g/kg) / Sea water Salinity (g/kg)	1.5	-
Gain Output Ratio	Distillate Flow / (Steam Condensate Flow Rate + De-superheating Flow)	8.65	-

Table 4.2: Secondary calculations

No.	Distillate Production (MIGPD)	Summer Operation		Winter Operation	
		6.0	7.2	6.0	7.2
1	SEAWATER SUPPLY TEMPERATURE °C	32.22	32.22	14.40	14.40
2	SEAWATER FLOW tonne/h.	10830.00	10166.00	4820.00	5095.00
3	VENT CONDENSER FLOW tonne/h.	1100.00	1240.00	1100.00	1240.00
4	RETURN SEA WATER FLOW tonne/h.	0.00	0.00	2904.30	2835.00
5	SEA WATER FLOW TO HEAT REJECTION SECECTION tonne/h.	9629.30	8826.00	6520.90	6591.00
6	SEA WATER INLET TEMPERATURETO HEAT REJECTION °C	32.22	32.20	23.90	23.90
7	BRINE RECIRCULATING FLOW tonne/h.	14286.00	12515.00	13663.00	11931.00
8	BRINE BLOW DOWN FLOW tonne/h.	1797.70	1987.00	1779.50	1959.00
9	LAST STAGE BRINE TEMPERATURE°C	40.50	41.21	35.18	35.18
10	MAKE UP WATER FLOW tonne/h.	2925.40	3340.00	2907.20	3312.00
11	BRINE INLET TO BRINE HEATER TEMPERATURE °C	84.89	103.02	82.13	100.59
12	BRINE OUTLET FROM BRINE HEATER TEMPERATURE °C	90.56	110.00	88.08	107.94
13	L.P. STEAM FLOW TO DISTILLATE tonne/h.	116.44	131.02	116.32	129.78
14	ATOMIZING STEAM FLOW TO DESUPER HEATER tonne/h.	5.66	5.66	5.66	5.66
15	DESUPER HEATER WATER FLOW tonne/h.	18.86	19.77	18.88	21.00
16	BRINE HEATER SUPPLY STEAM tonne/h.	140.96	156.45	140.96	156.45
17	EJECTOR STEAM SUPPLY tonne/h.	3.52	3.52	3.52	3.52
18	DISTILLATE FLOW tonne/h.	1127.70	1353.30	1127.70	1353.30
19	DISTILLATE TEMPERATURE °C	41.55	39.30	36.13	32.80
20	CONDENSATE RETURN FLOW tonne/h.	122.10	136.68	121.98	135.45
21	CONDENSATE TEMPERATURE °C	100.00	119.85	97.59	117.78
22	CIRCULATION RATIO	12.67	9.25	12.12	8.82
23	PERFORMANCE RATIO	8.00	8.85	8.00	8.65
24	FLASHING TEMPERATURE RANGE °C	50.06	68.79	52.69	72.76
25	TOTAL TEMPERATURE RANGE °C	58.34	77.78	73.64	93.50

Table4.3: Operating conditions of MSF distillers at Doha West Power Station (37).

4.4 Instrumentation Selection and Calibration

The proper operation of the distiller was most crucial during the evaluation of antiscalants. The precise measurements of the flows, temperatures, pressures and conductivity were essential in the daily monitoring of the plant performance.

Slight deviations would be reflected in the calculations of the fouling factors of the distiller and would affect the performance evaluation.

The unit instrumentation was classified into two kinds. The first kind consists of all instrumentation located in the central control room and the second kind relates to those which are sited locally. Though ideally data from both sources should be utilized it was decided, due to the volume of work required in the time available, to depend only on the data generated by the instruments in the control room.

Thermocouples were placed in strategically located thermo-wells as shown in Figures 4.2 and 4.3. Thus, it was necessary to carry out thorough cleaning of both inner and outer thermo-wells surfaces. The only other local instrumentation used consisted of simple local manometers placed in order to monitor the vacuum in the heater shell.

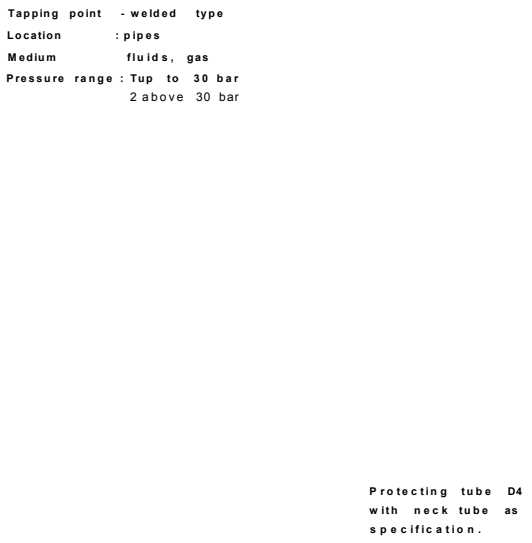


Figure 4.2: Thermo-couple and its thermo-well for measurement instrumentation (38)

Figure 4.3: Photo of the thermo-couple and its thermo-well for measurement instrumentation (39).

All instrumentation in the control room was calibrated according to the practice followed by the station. In addition, instrumentation related to temperature were also calibrated based on readings obtained using thermocouples inserted in the above mentioned thermo-wells.

4.5 Instrument Procedures Calibration

4.5.1 Control Room Instruments

A. Electronic Recorders

- Carry out technical cleaning of interior and exterior using approved cleaning spray, cleaning brushes and materials.
- Check gears and bearings, replace defective parts.
- Clean and lubricate all moving parts using appropriate lubricants.
- Clean and lubricate all moving parts using appropriate lubricants.
- Check all connecting wire terminals for tightness.
- Clean ink tubes, ink tanks and pens using the special cleaning wire.
- Re-fill the inking system and fit new chart.
- Calibrate and check writing system adjustment, carry out zero and span calibration check, record as found calibration details, adjust if necessary, and record final calibration details.

B. Pneumatic Recorders

- Clean air filters, air restrictor, and nozzle and flapper assemblies by using approved cleaning spray or cleaning wire.
- Check all of the internal pneumatic tubing for restrictions and foreign matter; clean/renew as found necessary.
- Check the operation of the recording mechanism and servo-mechanism.

- Clean and lubricate all moving parts using appropriate lubricants.
- Check all connecting wire terminals and pipe works for tightness and leaks.
- Clean ink tubes, ink tanks and pens using the special cleaning wire.
- Refill the inking system and fit new chart.
- Check writing system adjustment (pen contact pressure and ink tank).
- Carry out zero and span calibration check, record as found calibration details. Adjust if necessary, record final calibration details. All calibration records shall be retained and submitted for validation.
- Carry out a scale tracing check (0, 25, 50, 75 and 100%for upscale and downscale travel).

C. Indicators

a. Electrical

- Carry out technical cleaning of interior and exterior using approved cleaning spray, cleaning brushes and materials. Replace broken and clean indicator glass as found necessary.
- Carry out a calibration check at (0, 25, 50, 75 and 100%FSD). Rectify/replace as found necessary.
- Check all connecting wire terminals for tightness.

b. Pneumatic

- Carry out technical cleaning of interior and exterior using approved cleaning spray, cleaning brushes and materials. Replace broken and clean indicator glass as found necessary.
- Carry out a calibration check at (0, 25, 50, 75 and 100% FSD). Rectify/replace as found necessary.
- Check all pipe works for leaks.

D. Hand/Auto Stations, (Set Point Adjuster and Push Button Switch)

a. Electrical

- Carry out technical cleaning of interior, exterior and contacts, using approved cleaning spray, cleaning brushes and materials.
- Examine internal parts; replace defective parts as found necessary.
- Clean the control signal indicators replace broken and clean indicator glass, as found necessary. Carefully clean the control signal indicator scales, re-mark scales as found necessary to the original format

  and quality. Function check the push button switches, replace when required.
- Check lamps, replace if necessary.
- Check all connecting wire terminals for tightness.

b. Pneumatic

- Carry out technical cleaning of interior, exterior and contacts, using approved cleaning spray, cleaning brushes and materials.

  Examine internal parts; replace defective parts as found necessary.
- Clean the control signal indicators replace broken and clean indicator glass, as found necessary. Carefully clean the control signal indicator scales, re-mark scales as found necessary to the original format and quality. Function check the push button switches, replace when required.
- Check all pipe-work for leaks.

c. Counters (Integrators)

- Carry out technical cleaning of interior and exterior using approved cleaning spray, cleaning brushes and materials.
- Check gears and bearings, replace defective parts.
- Clean and lubricate all moving parts using appropriate lubricants.
- Check all connecting wire terminals for tightness.

d. Control Desk and Panel

- Carry out technical cleaning of interior and exterior using approved cleaning spray, cleaning brushes and materials.
- The cleaning shall be done before the unit start up, but shall be after the completion of the maintenance work and desk/panel painting.
- Check all connecting wire terminals for tightness, including the alarm/status windows inside the desk and panel.
- Replace all defective lamps associated with the alarm/status windows, desk/annunciation system.
- Check power points, lighting, telephone jacks, ventilation grills, glass panels and rubber beading, repair/replace as found necessary.

4.5.2 Transmitters

A. Electronic Transmitters

- Carry out transmitter inspection, clean exterior parts using approved cleaning spray, cleaning brushes and materials.
- Carry out technical cleaning of interior parts using approved cleaning spray, cleaning brushes and materials, if appropriate for the type of transmitter.

- Check transmitter operation, carry out a calibration check.
- Check root valves, isolating valves, equalizing valves, drain valves and vent valves. Service, repair or replace as found necessary.
- Check impulse lines and fittings, service, repair or replace as found necessary.
- Purge impulse pipe works before start-up and after line pressurization.
- Check output current and carry out leak check.

B. Pneumatic Transmitters

- Carry out transmitter inspection, clean exterior parts using approved cleaning spray, cleaning brushes and materials.
- Carry out technical cleaning of interior parts using approved cleaning spray, cleaning brushes and materials, if appropriate for the type of transmitter.
- Clean air filters, air restrictor, nozzle reducer and reducer tube assemblies by using appropriate cleaning spray or cleaning wire.
- Carry out the complete servicing of pneumatic relays, bushes and pressure gauges. The bushes shall be replaced if necessary. Calibrate the positioned pressure gauges.
- Check all of the internal pneumatic tubing for restrictions and foreign matter; clean/renew as found necessary.
- Clean and lubricate all moving parts using appropriate lubricants.
- Check all connecting pipe works for tightness and leaks.
- Check transmitter operation; carry out a calibration check.
- Check root valves, isolating valves, equalizing valves, drain valves and vent valves. Service, repair or replace as found necessary.
- Check impulse lines and fittings, service, repair or replace as found necessary.
- Purge impulse pipe works before start-up and after line pressurization.
- Check output current, carry out leak check.

4.5.3 Conductivity Measuring System

Designed to monitor the conductivities of demineralised water, steam condensate, water distillates, raw water and de-ionized water.

Procedure

- Thoroughly clean the electrodes with a nylon bottle brush using a warm water and detergent solution, without damaging the electrodes. Rinse with distilled water after cleaning.
- Inspect the analyzer assembly electrical connections, including the cell to transmitter wiring. Check complete external indicator/recorder circuit.
- Carry out calibration/function checks according to the manufacturer's technical specifications and procedures and check the transmitter alarm/trip set points, adjust if necessary.
- Carry out routine periodic maintenance according to the manufacturer's technical specifications and procedures.
- A system inspection, including calibration, shall be carried out weekly to ensure correct operation.

4.5.4 Local Gauges

All gauges serving all applications shall be serviced and calibrated. All associated valves and pipe works shall also be checked and repaired/replaced as necessary. Calibrations procedure as following:

A. Pressure, Differential Pressure, Filled Capillary Pressure and Level

- Remove from plant, clean and visually inspect external parts.

- Check the calibration using a certified dead-weight tester for rising and falling pressures, record results.
- If the calibration is in error, unacceptable hysteretic or other mechanical sticking/damaged, strip down the gauge, clean and visually inspect the internal parts, lubricate moving parts using approved lubricants. Repair as found necessary.
- Calibrate the gauge using a certified deadweight tester for rising and falling pressures, record results.
- Replace to plant.
- Check root valves, isolating valves, equalizing valves, drain valves and vent valves and manifold valves. Service, repair or replace as found necessary.
- Check impulse lines and fittings, service, repair or replace as found necessary.
- Purge impulse pipe works before start-up and after line pressurization.
- Check fittings for leakage and tighten if necessary.

B. Temperature (Solid Stem and Capillary)

- Remove from plant, clean and visually inspect external parts. Check the calibration, using approved standard test equipment, for rising and falling temperatures, record results.
- If the calibration is in error, shows unacceptable hysteretic or other mechanical sticking/damage, strip down the gauge, clean and visually inspect the internal parts, lubricate moving parts using approved lubricants and repair as found necessary.
- Calibrate the gauge, using approved standard test equipment, for rising and falling temperatures, record results.
- Replace to plant.
- Clean and check the sensing point thermo-well/pocket for mechanical damage and rust. Repair/replace as found necessary. Ensure the pocket is internally free from the ingress of foreign matter. Ensure that the sensing element is correctly fitted.

4.5.5 Switches

All switched serving all applications shall be serviced and calibrated. All associated valves and pipe works shall also be checked and repaired/replaced as necessary. Calibrations procedure as following:

A. Flow Switches

- Remove from plant, clean and visually inspect external parts.
- Clean and visually inspect internal parts, repair as found necessary.
- Check the switch operation.
- Check the set point of the switch using a workshop flow proved rig.
- Check root valves, isolating valves, equalizing valves, drain valves and vent valves and manifold valves. Service, repair or replace as found necessary.
- Check impulse lines and fittings, service, repair or replace as found necessary.
- Purge impulse pipe work before start-up and after line pressurization.
- Carry out circuit function check.
- Check fittings for leakage and tighten if necessary.

B. Pressure and Level Switches

- Remove from plant, clean and visually inspect external parts.
- Clean and check the level switch float and chamber. Float should be free from any minute formation of rust, or any holes that may allow the process medium to enter the float. Weld / replace as found necessary. Ensure that any weld repairs do not affect the operation of the float switch.

- Check the calibration, using a comparator and standard certified test gauge, for rising and falling pressures, record results.
- If the calibration is in error, unacceptable hysteretic or other mechanical sticking/damage, strip down the switch, clean and visually inspect the internal parts, lubricate moving parts using approved lubricants. Repair as found necessary.
- Calibrate the switch using a comparator and standard certified test gauge, for rising and falling pressures, record results.
- Replace to plant.
- Check root valves, isolating valves, equalizing valves, drain valves and vent valves and manifold valves. Service, repair or replace as found necessary.
- Check impulse lines and fittings, service, repair or replace as found necessary.
- Purge impulse pipe works before start-up and after line pressurization.
- Carry out circuit function check.
- Check fittings for leakage and tighten if necessary.

C. Temperature Switches

- Remove from plant, clean and visually inspect external parts. Long capillary switches may be maintained on the plant, if the appropriate portable test equipment is available.
- Check the calibration, using approved standard test equipment, for rising and falling temperatures, record results.
- If the calibration is in error, shows unacceptable hysteretic or other mechanical sticking/damage, strip down the gauge, clean and visually inspect the internal parts, lubricate moving parts using approved lubricants and repair as found necessary.
- Calibrate the gauge, using approved standard test equipment, for rising and falling temperatures, record results.

- Clean and check the sensing point thermo-well/pocket for mechanical damage and rust. Repair/replace as found necessary. Ensure the pocket is internally free from the ingress of foreign matter. Ensure that the sensing element is correctly fitted.
- Replace to plant.
- Carry out circuit function check.

D. Limit Switches

- Clean and visually inspect internal parts, repair as found necessary.
- Flexible conduit and cable glands to be checked and replaced as required. Damaged Glands to be replaced.

4.5.6 Temperature Instrumentation

Calibrations procedure as following:

A. Thermocouple Sensors & Thermo-wells

- Remove temperature sensor, clean and check the thermo-well/pocket, head, extension, porcelain base and terminals for mechanical damage and rust. Repair/replace as found necessary. Ensure the pocket is internally free from the ingress of foreign matter. Ensure that the sensing element is correctly fitted.
- Ensure that the signal cable grounds are securely connected to the earth.
- Repair/replace all damaged thermocouple cables, sleeves and their screens, as required.
- Check the thermocouple junction boxes, cold junction units and compensating cable, repair/replace as required.
- Carry out complete wiring loop function check for each unit.

B. Thermocouple Amplifiers

The following procedure shall be added for the temperature amplifiers, where fitted:

- Carry out calibration checks using an approved precision decade resistance box (0.1% accuracy or better) to simulate the temperature sensor.
- Carry out calibration checks on thermocouple amplifiers using an approved precision workshop potentiometer to simulate the temperature sensor.
- Carry out complete wiring loop function check for each circuit.

C. Temperature Monitoring Systems

The following procedure shall be added for the temperature monitoring systems:

- Carry out calibration checks using an approved precision decade resistance box (0.1% accuracy or better) to stimulate the temperature sensor.
- Carry out calibration checks on thermocouple instrumentation using an approved precision workshop potentiometer to simulate the temperature sensor.
- Use appropriate standard test equipment to measure the output.
- Check the temperature monitoring system alarm set point values, adjust if necessary.
- Check the operation of the multi-point switch unit. Clean the switch contacts using approved solvents. Repair and replace faulty items as found necessary.

4.6 Operational Troubles

By careful advance planning and good maintenance work, continuous operation of the distillation unit during the trial period was achieved. A few problems (such as: non-flowing of chemical additives, electrical motor and pump shaft problems, adjustment of the stroke of the pump, and mechanical pump impeller problems) did arise but these were sorted out quickly without any serious detrimental effect on the unit operation. The dosing pumps were closely monitored and cleaned to ensure the correct dosing rates were maintained.

The most serious problem that occurred concerned the steam supply to the brine heater. During normal operation the heater shell pressure stayed in vacuum, well within design value. However, on one particular day the pressure of the heater shell suddenly moved into positive pressure whereby the vacuum disappeared. This was very alarming as it could have been due to substantial scale formation inside the tubes of the brine heater.

The valve on the vent line connecting the brine heater to stage no.1 was opened, which lead to the gradual return of the vacuum in the brine heater shell. This proved that the loss of the vacuum was not a consequence of scale formation but rather to ingress of steam due to the existence of a hole in the steam line section under vacuum downstream of the de-super-heater.

The most likely location of the hole was the expansion joint at one of steam line branches feeding steam to the brine heater. As the opening of the vent valve lead to the overloading of the air ejectors and thus disturbance of the whole unit, it was decided to close the valve and accept the existence of positive pressure in the shell of the brine heater once the reason for it has been identified.

During the above investigation many parameters were disturbed. However, once the vent was kept in a closed position and after the ingress of air stopped due to equalization between the pressure in the steam line and the atmosphere, most parameters stabilized with the exception of the heat transfer coefficient and the fouling factor of the heat input section. Both were affected due to reduced heat transfer as a result of the presence of air in the brine heater shell.

4.7 Pre and Post Trial Inspections

The pre and post-trial inspection is one of the main criteria in ascertaining the performance of the antiscalant. The pre-trial inspection serves as the baseline of the actual condition of the tubes especially the sections of the brine heater inlet and outlet tubes which are the highest temperature zones of the evaporator and predominantly these are prone to scaling that will easily adhere if the scale inhibition property of the antiscalant is not effective; and its procedure as following:

4.7.1 Flash Chambers

- Scale covering places and its thickness to be checked as a whole in the heat recovery section and heat rejection section.
- Accumulated scale flakes on the floor, near the weir boxes or inside and at the corners must be removed.
- Corroded areas must have surface preparation followed by patch painting.
- Demisters condition at the upper and lower decks must be checked if acid cleaning is needed or not and repair of any damaged frames / meshes must be done before re-fitting.
- Sacrificial anodes must be cleaned and consumed anodes to be replaced.
- Pressure, temperature and conductivity thermocouples condition must be checked if they are covered with a scale layer or clean.

4.7.2 De-aerator

- Condition of plastic packing pieces must be checked and repairing of any damaged metal net plates; this must be ensured to avoid any escape of the plastic packing pieces through the gaps.

4.7.3 Water Boxes

- Accumulated flake scale on the floor must be checked that may block the bottom tubes, tubes condition must be checked whether they are

covered with scale or not and point out the scale places and scale thickness, where it is increasing and diminishing. Boxes that needs hydro-jetting to clean the condenser tubes, it is better to do all preferably.

- Corroded area must be pointed out.
- Heat rejection water boxes surface must be checked for any accumulated salt (should be removed by flushing).

4.7.4 Brine heater

- Tubes condition must be checked and pointed out the places of any scale inside the tubes and its thickness to be measured and recorded.
- Condition must be checked if there are any scale or not and its thickness must be pointed out with attaching photos.

4.7.5 Ball Cleaning System

- Mesh strainer must be checked, it may be blocked with scale flakes and sponge balls.
- Sponge balls condition to be checked and if any are missing.

4.8 Data Analysis and Evaluation

The 90 days evaluation of antiscalants with measured dosing rates give a good indication of how the antiscalant is able to inhibit scale formation at the given top brine temperatures between 105 - 110 C.

The correct operation of the ball cleaning distillers is extremely important during the evaluation of the antiscalant and the physical monitoring of the operation in the control room and on the plant is essential to ensure that the balls are circulated in the tubes in the Brine Heater and Heat Recovery Section water boxes.

The antiscalant is dosed in neat form and dosing flows are automatically set in the control room.

It was decided that distiller performance monitoring should be performed daily and on a continuous basis. This gave a clear insight into factors affecting the overall unit operation and the opportunity to vary any parameter based on the analyzed data of the previous day. It also provided a very early warning of any instrument malfunction or unit performance decline. Performance parameters were produced in a graphic form on a daily basis for the duration of the trial.

4.9 Operational Parameters

Operational parameters were monitored for two reasons. The first is to ensure that unit operation is kept within the design constraints. The second reason is to enable checking of the accuracy of some of the collected input data. The monitored operational parameters include the following:

- Re-circulating Brine Flow-rate
- Seawater Flow-rate
- Recirculation Ratio
- Make-up/Distillate Flow Ratio
- Re-circulating Brine/Cooling Seawater Flow Ratio
- Top Brine Temperature
- Bottom Brine Temperature
- Inlet Seawater Temperature
- Flashing Range
- Total Temperature Range
- Concentration Ratio

The logarithmic mean temperature difference and the brine tube velocity were both determined for each unit section: namely, heat input, heat gain and heat rejection sections. In the case of some stream flows and concentration factors, both observed and calculated values were obtained.

4.10 Performance Parameters

It is important to remark that the effectiveness of the chemical was determined through monitoring the overall performance of the unit. The performance parameters determined include:

- Distillate Flow-rate
- Steam Flow-rate
- Gain Output Ratio
- Performance Ratio
- Steam Consumption
- Heat Energy Transfer
- Heat Transfer Coefficient
- Fouling Factor

The performance parameters, including heat energy transfers, heat transfer coefficients and fouling factors, were all determined for the three sections of the unit. In addition both observed and calculated gain output ratios were obtained.

Literature studies on the MSF desalination plants and its antiscalants provide information on thermal coefficients:

- Most of the data are presented in a form of single values for the overall heat transfer coefficients for the brine heater and heat recovery section. These coefficients include the effect of fouling resistance.
- Optimization of the MSF plants requires relationships of the effects of the operating variables on thermodynamic efficiency.

The main design features of MSF units are:

- All units are cross-flow type with brine recirculation;
- The flashing stages are arranged in two tiers;
- The rated unit capacity varies between 25,000 and 32,000 m3/day;
- The number of flashing stages is 21-23 in the heat gain section andthree in the heat rejection section; this gives a total number of stages of 24-26;
- The heat transfer area varies from 77,206 to 77,441 m2 in the heat gain section and from 9444 to 10,194 m2 in the heat rejection section;
- The outer diameter of the tubes in the heat gain section is 43 mm and in the heat rejection section is 33.4 mm.

The test data include the following cases:

- Performance test data, which were collected over a period of 7-24 h;
- Reliability test data, which were obtained from clean units;
- Minimum load test data, which were collected account for 70-75% of the normal design load;
- Overload test data, which were collected account for 110-115% of the normal design load;
- High temperature operation;

The ranges of the operating variables were:

- Top brine temperature = 91.1-106.55°C;
- Brine blow-down temperature = 30.5-40.6°C;
- Brine velocity inside the tubes = 1.32-2.34m s'1;
- Performance ratio = 8.37-10.55.

4. 1 Instrumentation and Error Analysis.

Determination of heat transfer coefficient required measurements of the following system variables (20):

- Temperatures of inlet/outlet brine flowing inside the condenser tubes for each flashing stage in the heat recovery section;
- Temperatures of inlet/outlet intake seawater flowing inside the condenser tubes for each flashing stage in the heat rejection;
- Temperature of the condensing vapour in each flashing chamber;
- Temperature of the brine stream leaving each flashing stage;
- Height of the brine pool in each flashing stage;
- Brine flow rate between the flashing stage;
- Brine recycle flow rate inside the condenser tubes;
- Seawater flow rate inside the condenser tubes in the heat recovery section.

Measurements of the above parameters utilized the following instrumentation:

- Thermocouples were used to measure the temperatures of the flashing brine, vapour above the demister, and the seawater/brine flowing inside the condenser tubes. The thermocouple probes were (K) type with a measuring accuracy of ±0.5°C.
- Level gauges were used to measure the brine height inside the flashing stages. The gauge accuracy was ±0.01 m.
- An orifice/manometer combination was used for measurements of the flow rate. The device accuracy was rated over a range of 0.01m of water in the manometer readings. This gave a flow rate error of 1% in the high temperature section and 7% in the low temperature section.

The number of condenser tubes, the outside diameter of the tubes and the tube length were used to determine the heat transfer area in each section.

4.12 Operating Conditions

The comparative studies of the three different antiscalants were carried out under same conditions of temperatures antiscalant dosing rates and concentration ratios. All antiscalant were evaluated at a TBT of 105°C and a dose rate of 1.5.

Chapter 5

Comparison Study

Comparison Study

5.1 Introduction

This chapter summarizes the evaluated thermal performance results of applying three different antiscalant to an MSF plant. The aim of the trial was to perform the tests under comparable operation conditions in order to evaluate the performance of the plant and the efficacy of the three antiscalants under test.

The behaviour of major operational and performance parameters over the trial periods will be comprehensively compared and presented in a graphic forms as a function of elapsed time (days) for the three different antiscalants. Table 5.1, 5.2 and 5.3 lists and summarises the Average, Maximum, and Minimum, Design value, % difference, Standard deviation and the Coefficient of variation values of the important operational parameters. These values are analysed to indicate where there have been fluctuations from the design parameters, the reasons for the fluctuations and the affect it could have on the validity of the comparative studies.

The % difference refers to the difference between the average and the design value. The standard deviation is for the whole population (StdevP). The Coefficient of variation is given by the standard deviation^-mean and gives an indication in the degree of fluctuation of the values over the test period.

The full spreadsheet analysis for the 90 day data is provided on a CD rom in Appendix G.

Most of the parameters were direct readings but the derived data were:

Flash range:

Brine Fleater Outlet Temperature - Last Stage Brine Temperature

Total Temperature Range:

Brine Heater Outlet Temperature - SeaWater to Distiller Temperature

Circulation Ratio:

Brine Recirculation Flowrate ÷ Distillate Flowrate

Gain Output Ratio:

Distillate Flowrate ÷ (Return Condensate Flowrate + Spray Water Flowrate)

Overall Heat Transfer Coefficient (BH) Equation 3.1

Overall Heat Transfer Coefficient (Rec.Sec.) Equation 3.4

Fouling Factor (BH) Equation 3.3

Fouling Factor (Rec. Sec.) Equation 3.5

Applying an error analysis to the data indicates small inaccuracies for most of the results. The largest inaccuracies will be for the calculation of the heat transfer coefficients due to the number of temperature differences used in the calculation. The thermocouple readings are ±0.5 K so each temperature difference could result in a ±1.0 K error due to instrumentation accuracy.

Using equation 3.4 and substituting in typical data indicates a % error due to the accuracy of the thermocouples as 5.3%.

Parameters	Unit	Ave	Max	Min	Design Value	% Difference	StdevP	Coefficient of variation%
Seawater Inlet Flow rate	Tonne/h	7876	8644	6725	5095	55	548.8	7.0
Seawater Make-up Flow rate	Tonne/h	3439	3578	3267	3312	4	78.6	2.3
Brine Recirculation Flow rate	Tonne/h	12161.	12286	12047	11931	2	67.1	0.6
Brine Blow-down Flow rate	Tonne/h	2049	2212	1730	1959	5	114.6	5.6
Distillate Flow rate	Tonne/h	1409	1449	1363	1353	4	23.2	1.6
Brine Heater Shell Pressure	bar(a)	1.29	1.45	1.20	1.50	-14	0.045	3.5
Brine Heater Inlet Temperature	°C	97.4	99.4	94.5	100.6	-3	1.34	1.4
Brine Heater Outlet Temperature	°C	103.4	106.5	100.5	107.9	-4	1.55	1.5
Condensate Temperature	°C	111.2	115.0	106.9	117.8	-6	2.53	2.3
LP Steam to Brine Heater Temperature	°C	110.9	114.4	108.5	117.8	-6	1.36	1.2
Sea Water to Distiller Temperature	°C	21.5	26.0	20.0	23.9	-10	1.98	9.2
First Stage Brine Temperature	°C	95.6	99.5	91.9	100.6	-5	1.32	1.4
Last Stage Brine Temperature	°C	35.5	36.7	33.9	35.1	1	0.93	2.6
Flash Range	K	67.8	72.5	65.0	72.7	-7	1.83	2.7
Total Temperature Range	K	81.8	86.5	74.5	93.5	-12	3.32	4.1
Circulation Ratio		8.63	8.95	8.41	8.82	-2	0.154	1.8
Distillate Conductivity	µs/cm	6.63	9.70	3.60	60.00	-89	1.219	18.4
Gain Output Ratio (GOR)		9.90	10.58	9.44	8.65	14	0.246	2.5
Chloride Ratio		1.41	1.49	1.29	1.50	-6	0.038	2.7
Antiscalant Dosing Rate	ppm	1.46	1.54	1.40	1.50	-27	0.034	2.4
Overall Heat Transfer Coefficient (BH)	kW/m^2K	2.18	2.97	1.46	2.060	6	0.286	13.1
Overall Heat Transfer Coefficient (Rec. Sec.)	kW/m^2K	2.33	3.84	1.50	2.73	-15	0.523	22.5
Fouling Factor, B/H	m^2K/kW	0.24	0.46	0.11	0.264	-7	0.064	26.3
Fouling Factor Heat Rec. Sec.	m^2K/kW	0.23	0.45	0.04	0.15	56	0.097	41.5

Table5.1: Mean and Standard deviations values of antiscalant A

Parameters	Unit	Ave	Max	Min	Design Value	% Difference	StdevP	Coefficient of variation%
Seawater Inlet Flow rate	Tonne/h	9255	9330	9143	5095	82	54.8	0.6
Seawater Make-up Flow rate	Tonne/h	3535	3561	3498	3312	7	22.1	0.6
Brine Recirculation Flow rate	Tonne/h	12686	12853	12304	11931	6	149.5	1.2
Brine Blow-down Flow rate	Tonne/h	2101	2228	2038	1959	7	40.8	1.9
Distillate Flow rate	Tonne/h	1361	1384	1335	1353	1	12.2	0.9
Brine Heater Shell Pressure	bar(a)	1.31	1.40	1.20	1.5	-13	0.064	4.9
Brine Heater Inlet Temperature	°C	97.0	98.1	96.4	100.6	-4	0.38	0.4
Brine Heater Outlet Temperature	°C	104.4	107.4	102.0	107.9	-3	0.74	0.7
Condensate Temperature	°C	111.6	113.0	109.8	117.8	-5	0.61	0.5
LP Steam to Brine Heater Temperature	°C	111.1	114.4	108.4	117.8	-6	1.34	1.2
Sea Water to Distiller Temperature	°C	23.3	25.0	22.0	23.9	-2	1.13	4.8
First Stage Brine Temperature	°C	95.7	99.3	92.9	100.6	-5	1.20	1.3
Last Stage Brine Temperature	°C	32.5	36.0	30.0	35.2	-8	1.20	3.7
Flash Range	K	71.9	73.7	70.1	72.8	-1	0.79	1.1
Total Temperature Range	K	80.9	85.4	77.4	93.5	-13	1.39	1.7
Circulation Ratio		9.39	9.45	9.30	8.82	6	0.048	0.5
Distillate Conductivity	µs/cm	3.26	7.00	1.20	60	-95	1.481	45.4
Gain Output Ratio (GOR)		8.64	8.88	8.36	8.65	0	0.110	1.3
Chloride Ratio		1.37	1.45	1.23	1.5	-8	0.041	3.0
Antiscalant Dosing Rate	ppm	1.45	1.60	1.33	1.50	-3	0.070	4.8
Overall Heat Transfer Coefficient (BH)	kW/m^2K	2.80	4.49	1.64	2.060	36	0.475	17.0
Overall Heat Transfer Coefficient (Rec. Sec.)	kW/m^2K	1.84	2.89	1.11	2.728	-33	0.363	19.7
Fouling Factor, B/H	m^2K/kW	0.15	0.39	0.00	0.264	-45	0.062	42.5
Fouling Factor Heat Rec. Sec.	m^2K/kW	0.35	0.69	0.13	0.150	132	0.103	29.8

Table5.2: Mean and Standard deviations values of antiscalant B

Parameters	Unit	Ave	Max	Min	Design Value	% Difference	StdevP	Coefficient of variation%
Seawater Inlet Flow rate	Tonne/h	7635	8563	6769	5095	50	391.3	5.1
Seawater Make-up Flow rate	Tonne/h	3459	3575	3340	3312	4	62.4	1.8
Brine Recirculation Flow rate	Tonne/h	11243	12762	10184	11931	-6	857.5	7.6
Brine Blow-down Flow rate	Tonne/h	1891.62	2031	1655	1959	-3	85.8	4.5
Distillate Flow rate	Tonne/h	1379.56	1412	1356	1353	2	13.4	1.0
Brine Heater Shell Pressure	bar(a)	1.34	1.40	1.30	1.5	-10	0.035	2.6
Brine Heater Inlet Temperature	°C	96.7	97.5	96.4	100.6	-4	0.30	0.3
Brine Heater Outlet Temperature	°C	103.7	103.9	103.7	107.9	-4	0.08	0.1
Condensate Temperature	°C	111.3	112.1	110.2	117.8	-5	0.46	0.4
LP Steam to Brine Heater Temperature	°C	112.1	114.0	110.0	117.8	-5	0.86	0.8
Sea Water to Distiller Temperature	°C	23.5	25.0	22.0	23.9	-2	1.08	4.6
First Stage Brine Temperature	°C	95.8	97.6	94.0	100.6	-5	0.89	0.9
Last Stage Brine Temperature	°C	35.8	38.7	32.7	35.2	2	1.36	3.8
Flash Range	K	67.9	71.2	65.0	72.8	-7	1.37	2.0
Total Temperature Range	K	80.2	81.9	78.7	93.5	-14	1.07	1.3
Circulation Ratio		8.11	9.40	7.48	8.82	-8	0.532	6.6
Distillate Conductivity	µs/cm	3.33	8.70	1.10	60	-94	2.103	63.2
Gain Output Ratio (GOR)		9.35	9.59	9.08	8.65	8	0.102	1.1
Chloride Ratio		1.43	1.50	1.30	1.5	-5	0.043	3.0
Antiscalant Dosing Rate	ppm	1.46	1.54	1.36	1.50	-3	0.034	2.4
Overall Heat Transfer Coefficient (BH)	kW/m^2K	2.29	2.80	1.91	2.060	11	0.203	8.9
Overall Heat Transfer Coefficient (Rec. Sec.)	kW/m^2K	1.85	2.74	1.16	2.728	-32	0.399	21.5
Fouling Factor, B/H	m^2K/kW	0.22	0.30	0.14	0.264	-17	0.038	17.3
Fouling Factor Heat Rec. Sec.	m^2K/kW	0.35	0.64	0.15	0.150	132	0.119	34.3

Table5.3: Mean and Standard deviations values of antiscalant C

The performance result parameters analysed include the following:

Figure 5.1: Measured Sea Water Inlet Flow Rate

The design sea water inlet flow of 5095 tonne/h is based on a sea water temperature that does not require tempering before entering the distillers as make-up water. During the winter, when the three trials were conducted, a sea water recirculation pump is used to temper the make-up water and an additional flow rate is necessary dependant on the sea water temperature and the amount of cooling water required for the heat rejection section. The higher flow rates, up to 90% above the design value, are used to balance the plant and are not an indication of poor control of the inlet flow rate.

Figure 5.2: Measured Sea Water Make-up Flow Rate

The data shown in figure 5.2 represent the variation of the measured sea water make-up flow rate with elapsed time (days).

Although the stated design sea water make-up flow rate for DWPS distillers for the winter mode is 3311.9tonnes/h as indicated in the plot, in practice, it was normally maintained around 3500 tonnes/h. The dosing regime for antiscalant, sodium sulphite and antifoam is based on the plant operational value of 3500 tonnes/h as shown by the dotted line in figure 5.2. The control of the plant for all three tests was good, the average values being within 2% of the plant operational value and the highest coefficient of variation was 2.3% for antiscalant A.

Incidences of the unit operating with high concentration ratio and consequent risk of precipitation of scales did not appear as any significant threat.

Increase in make-up flow lowers salt concentration in the brine stream, which reduces the specific mass of brine and the boiling point elevation, which in turn lowers the steam consumption and blow-down salt concentration. Consequently some improvement of gain output ratio will be expected. MSF plant normally operates at high make-up flow.

Figure 5.3: Measured Brine Recirculation Flow Rate

Figure 5.3 shows the variation of measured brine recirculation flow rate with elapsed time (days). For the operation with antiscalant C, the brine re-circulation flow was consistently significantly below (-10%) the design flow rate for about 70% of the test period. Importantly all three flows at the end of the test were close to the design value.

For antiscalant A operation, the Brine Recirculation flow rate has been maintained marginally above the design flow rate during entire test period (the average is above by 2%) and for antiscalant B operation, it is above the design value for the whole test period, the average being above by 6%.

Re-circulating brine Flow rate is a very important operational variable affecting the performance of the plant. Increase in brine re-circulation flow should bring about an increase in distillate production, while it causes a reduction in gain output ratio. This does not correlate with the results obtained for the three tests conducted which can be explained by variations in the other parameters.

The lower limit of brine recirculation flow rate is determined to avoid problem of scaling due to low brine velocity and the upper limit is decided to avoid high brine velocity that may cause erosion of tubes and carry-over of brine to the distillate. Usually the plant is run close to the upper limit. For all three tests the upper and lower limits were not exceeded.

Figure 5.4: Measured Distillate Flow Rate

The average distillate production rates for all three tests were higher than the design value (4%, 1% and 2% respectively) with antiscalant A outperforming B & C. The variance in the distillate production is due to changes in the operational parameters of the distiller rather than the effect of fouling due to scale deposition.

Some striking anomalies are noticed regarding different distillate production data under consideration. For identical plants, operating with similarly close process parameters, distillate production rate is likely to be affected mainly by the magnitude

of the brine recirculation flow rate and obtaining the correct flash range. Accordingly on this basis the expected distillate production rate should have been highest during the test period for antiscalant B operation. Surprisingly, it was found to be the lowest compared to operation with the other two antiscalants.

Figure 5.5: Measured Brine Heater Shell Pressure

It should be noted that LP steam supply to the de-super-heater changes in quality depending on the condition of the extracted steam from the turbine which is a function of the turbine loading and the subsequent pressure reduction exercised by the pressure reducing valves. The shell pressure is adjusted to ensure enough heat transfer, so under good working conditions the pressure is lower than the design pressure. When fouling occurs the shell pressure would be increased in order to achieve the necessary heat transfer rate.

Figure 5.5 shows that the Brine Heater Shell Pressure for all three tests was below the design value (the average values being below by 14, 13 and 10% respectively). This is an indication that fouling was not occurring. As for the trend it can be seen that the coefficient of variation was highest for antiscalant B (4.9%).

Figure 5.6: Measured Brine Heater Inlet Temperature

Figure 5.6 indicates that there is close control of the brine heater inlet temperature, especially for antiscalants B & C. The highest variations are for antiscalants A, but even this has a low coefficient of variation of only 1.4%. The average values are marginally below the design value (3, 4 & 4% respectively).

Figure 5.7: Measured Brine Heater Outlet Temperature (TBT)

One major parameter in assessing the performance evaluation of the three antiscalants is the brine heater outlet temperature or the top brine temperature (TBT) which plays a very important part with respect to the performance of MSF plants.

An increase of TBT improves the efficiency of the MSF distillation process (or, conversely, allows more economical evaporators with a given efficiency). An increase in TBT will increase scale deposition and hence the need for antiscalant treatments.

Conversely a high TBT can have negative effects and the designer has to examine carefully all the implications on plant availability and durability related to the selection of the TBT, and the consequent measures to take in terms of corrosion prevention and scale control.

Figure 5.7 shows a daily plot of measured brine heater outlet temperature. The design value is 107.9°C but for test operation with all three antiscalants a practical value of 105°C was chosen. Very close control was obtained for antiscalants B & C and for A the coefficient of variation was only 1.4%. The averages for all three antiscalants were within 2% of the controlled temperature of 105°C. It is important that all three tests had close control of this parameter as it known to have a major influence on plant performance.

Figure 5.8: Measured condensate temperature

The condensate is the heat source for the brine heater which can be seen from the high degree of correlation between figures 5.8 and 5.7. There is a corresponding fall in the condensate temperature antiscalant A to that of the Brine Heater temperature starting on day 39.

It can be seen from the plot that there is no increasing trend in condensate temperature in the course of test. It is almost the same and steady during the entire trial period. This indicates that there was no significant scaling of the brine heater tubes during the trial.

Figure 5.9: Calculated flash range

By definition, flash range is the difference between the top brine temperature and last stage brine temperature. The output of MSF distillation plants depends on the flash range. It therefore follows that if the top brine temperature is raised then significant improvements in distillate outputs and operating efficiency are possible. Thus, controlling the top brine temperature is an important aspect of the MSF process.

There are two ways in which the flash range can be increased. The first is by lowering the last stage brine temperature which is dependent on the seawater temperature. The other method involves raising the top temperature. At this end, there is need to consideration calcium sulphate scaling, especially at temperatures over 121 °C. At lower temperatures, scaling due to CaC_{03} and $Mg(OH)_2$ can be present, if not controlled properly, and as discussed earlier this does not happen in DWPS distillers, where the top brine temperature is controlled.

Antiscalant B had the highest flash range which was not due to a high TBT but to a marginally lower last stage brine temperature. There was close control of this parameter the spread of the average values being 5.7% and the highest coefficient of variation being only 2.7%

Figure 5.10: Calculated total temperature range

By definition, the total temperature range is the brine heater outlet temperature (TBT) minus the Sea water to distiller temperature which enters the tube side of last Heat Rejection Stage.

The tests were conducted in winter mode operation which resulted in a tempered sea water temperature. This is the reason for a higher design value of total temperature range compared to the test operation data for all three antiscalants. As, total temperature range is the difference of TBT and inlet temperature to the 24th stage tube side, the behavioural trend of change, in this regard is similar to the change observed in case of the Top Brine Temperature for all three antiscalants.

The spread for the average values is only 2% and the coefficient of variation was highest for antiscalants A (4.1%).

Figure 5.11: Calculated Circulation Ratio

By definition, Circulation Ratio is the ratio of brine recirculation flow to distillate production rate. The design value of brine recirculation flow is 11931 tonnes/h and design distillate production rate is 1353 tonnes/h and accordingly the design value of circulation ratio is 9.82.

From figure 5.11, it can observed that the circulation ratio was highest for antiscalant B, the average value being above the design value by 6%. By the end of the tests the circulation rates for A and C were marginally below the design value by 2%.

Circulation Ratio is an important parameter for monitoring the performance of the evaporator and the value should preferably be above the design value. As, lower circulation ratio values relate to lower brine recirculation flow rates this in turn results in reduced brine velocity in the brine heater and stage condenser tubes. Reduced brine recirculation flow rates may result in increased fouling.

Figure 5.12: Measured Distillate Conductivity

Guarantee value (maximum allowable) for distillate is 30 ppm T.D.S., which is equivalent to 60µs/cm. For all tests runs it never exceeded 10µs/cm. The conductivity of the distillate was maintained and there was no deterioration in the quality throughout the optimization trial periods.

This also indicates effective control rendered by the antifoam used. The higher brine recirculation flow (for antiscalants B) did not cause a higher brine level in the stages and consequent carry over, which would have resulted in higher levels of conductivity readings.

Figure 5.13: Calculated Gain Output Ratio

By definition Gain Output Ratio or GOR is the ratio of distillate produced and steam used. This is an important parameter for monitoring performance and efficiency of the evaporator. Along with brine heater shell pressure and condensate temperature data analysis, the extent of scaling of the brine heater tubes is assessed and judgment for the requirement of acid cleaning of any evaporator is made.

Antiscalants A and C out-performed the design value by 14% and 8% respectively while the average value for antiscalant B was the same as the design value.

The value of GOR is a compromise between production and economy. When higher distillate production is required this is achieved by increasing brine recirculation, which reduces GOR, but when economy is required the plant is run at a lower brine recirculation rate.

Figure 5.14: Measured Chloride Ratio chemically

The seawater at the intake is chlorinated to control marine bi-fouling. The chlorine ratio in the water is closely monitored to ensure it does not exceed environmental limits and to reduce the possibility of corrosion of the plant.

Only antiscalant C on three days marginally exceeded the design limit of 1.5 ppm. For the rest of the time the average values were kept below the limit by 6%, 8% and 5% for antiscalants A, B and C respectively.

The Chloride Ratio was well maintained to desired level during the trial. It shows that the distiller operated with proper seawater make up flow and correct dosing rate.

Figure 5.15: Calculated antiscalant dosing rate

To compare the effectiveness of the three antiscalants the dose rate was controlled to match or to be slightly lower than the new dose rate under test of 1.5 ppm. Figure

5.15 shows that the average dosing rates of the three antiscalants were marginally less than the dose rate under test by 3%.

For all three antiscalant test operations close control of TBT was maintained around 105°C. Also, in the course of the test operation of all three types of antiscalant, there had not been any significant increase of brine heater shell pressure, condensate temperature and GOR did not indicate any signs of the adverse effects of the antiscalants operating at this new dosing rate of 1.5 ppm. From this is can be considered that all three types of antiscalant performed well.

Figure 5.16: Calculated Overall Heat Transfer Coefficient (brine heater).

The heat transfer coefficient was monitored using the temperature readings in the brine heater and the steam condensate together with the brine re-circulation flow. In the MSF plant there is a requirement to evaporate water from a liquid phase and to condense the vapour to form the distillate. The overall heat transfer coefficient for the brine heater for all three antiscalaants was maintained, with antiscalant B giving the best value. It has been found to be consistently significantly better than the design value throughout the period, the average value being 82.5% higher. The corresponding values for antiscalants A and C being +49.2% and +67.2% respectively.

Some obtained values of the Brine Heater overall heat transfer coefficient were even higher than that for the clean tube overall heat transfer coefficient. This can be attributed to higher brine recirculation flow and the low condensate temperatures over the test period.

There was no downward trend in the heat transfer coefficient which indicates that there is no extensive fouling in the distiller and that the heat transfer surfaces were giving good heat transfer between the re-circulating brine stream and the condensing vapour.

Figure 5.17: Calculated Overall Heat Transfer Coefficient (Recovery Heat Section)

To calculate the overall heat transfer coefficient for the recovery section requires five temperature differences which give rise to the high coefficient of variation for this parameter, 14.2%, 10.3% and 11.0% for antiscalants A, B and C respectively. The averages show that antiscalant A had the best performance being above the design value by 19.0% compared with B and C being 9.0% and 9.9% above.

For antiscalant A test operation, brine recirculation flow rate was close to the design flow rate during the entire test period and also the LP steam flow rate to the brine heater did not vary very much.

The values obtained for all three antiscalants remained steady during the test period inferring that no significant scaling occurred in the Heat Recovery Section tubes during the trial period.

Figure 5.18: Calculated fouling factor of brine heater

The fouling factor is the ultimate parameter to be used when determining the cleanliness of heat transfer surfaces and, thus, is a measure of the effectiveness of antiscalant chemical treatments.

The daily average fouling factor was automatically calculated and used as the basis of strict monitoring of the plant performance. If for any reason fouling was taking place, the other indicating factors would be observed such as the production, performance ratio and the brine heater shell pressure.

Fouling factor has an inverse trend to that of the overall heat transfer coefficient. As such, from the plot, it is seen that the fouling factor is least for antiscalant B test operation.

The averages for the Brine Heater tube fouling factor are less than the design value by 59%, 81% and 73% respectively for antiscalants A, B and C. This was due to the good performance of antiscalants in conjunction with effective use of on-load sponge ball cleaning system to avoid tube scaling.

Figure 5.19: Calculated Fouling Factor Heat Recovery Section

Figure 5.19 compares the three calculated fouling factor of the heat recovery section. Fouling factor of the heat Recovery Section has an inverse trend to the overall heat transfer results. Antiscalant A has the best values of fouling factor among the three types of antiscalants. The averages values being better than the design value by 38%, 21% and 23% respectively.

Figure 5.1: Measured SeaWater Inlet Flowrate

Figure 5.2: Measured Make-up Flowrate

Figure 5.3: Measured Brine Re-circulation Flowrate

Figure 5.4: Measured Distillate Flowrate

Figure 5.5: Measured Brine Heater Shell Pressure

Figure 5.6: Measured Brine Heater Inlet Temperature

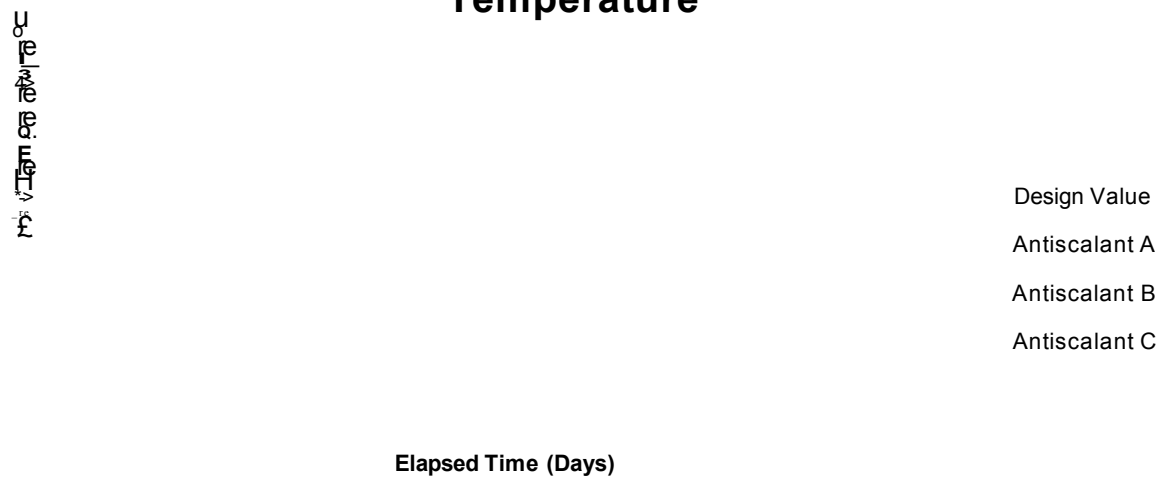


Figure 5.7: Measured Brine Heater Outlet Temperature

Figure 5.8: Measured Condensate Temperature

Figure 5.9: Calculated Flash Range

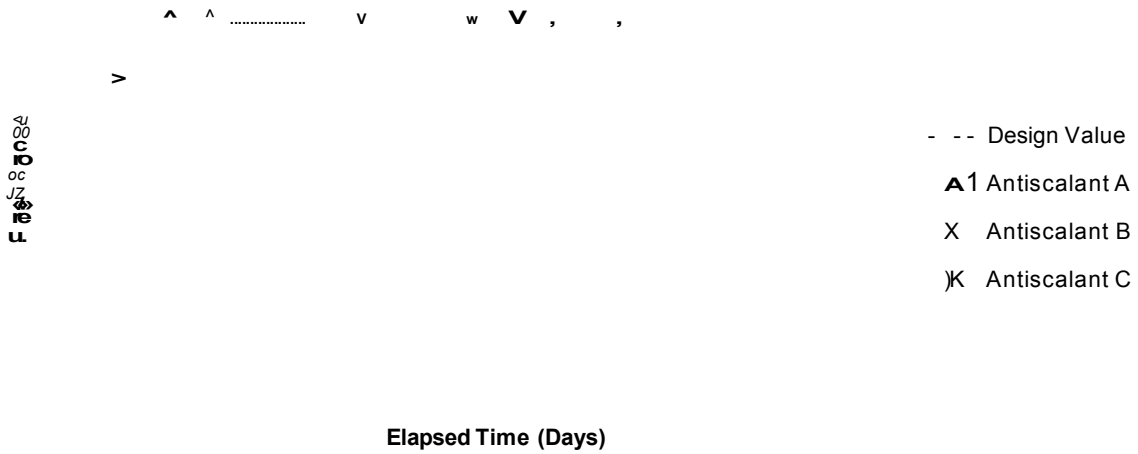
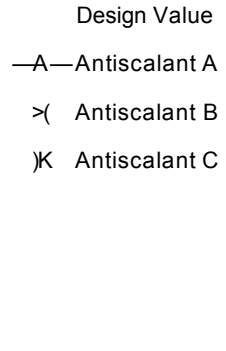
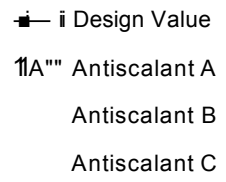


Figure 5.10: Calculated Total Temperature Range



Elapsed Time (Days)

Figure 5.11: Calculated Circulation Ratio



Elapsed Time (Days)

Figure 5.12: Measured Distillate Conductivity

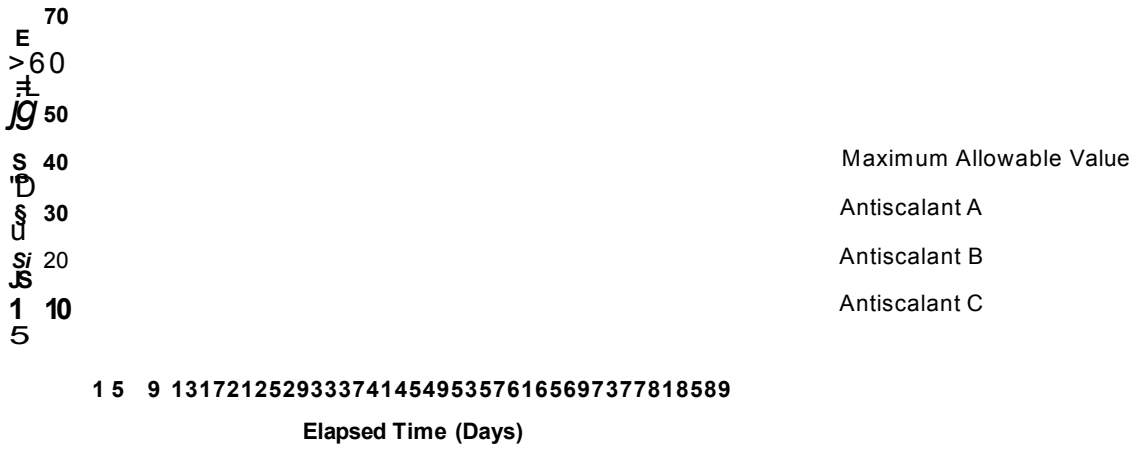


Figure 5.13: Calculated Gain Output Ratio (GOR)

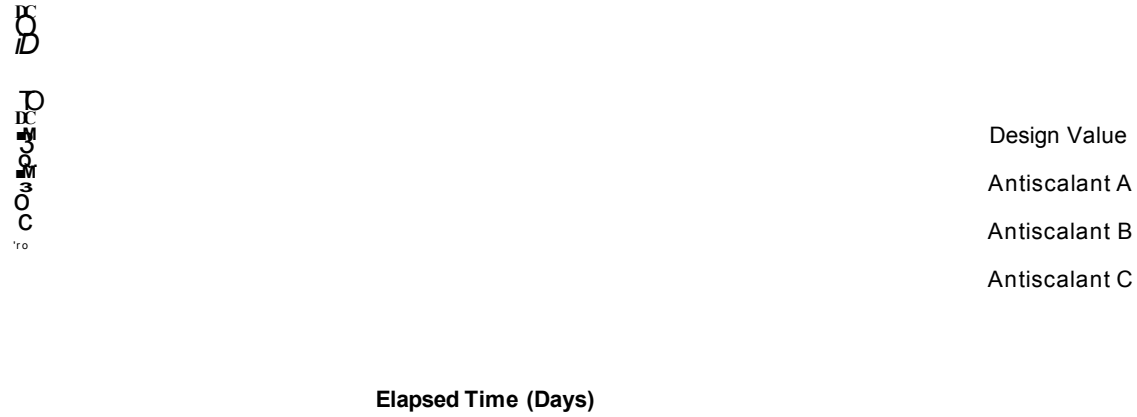


Figure 5.14: Measured Chloride Ratio Chemically

Figure 5.15: Calculated Antiscalant Dosing Rate

Figure 5.16: Calculated Heat Transfer Coefficient (BH)

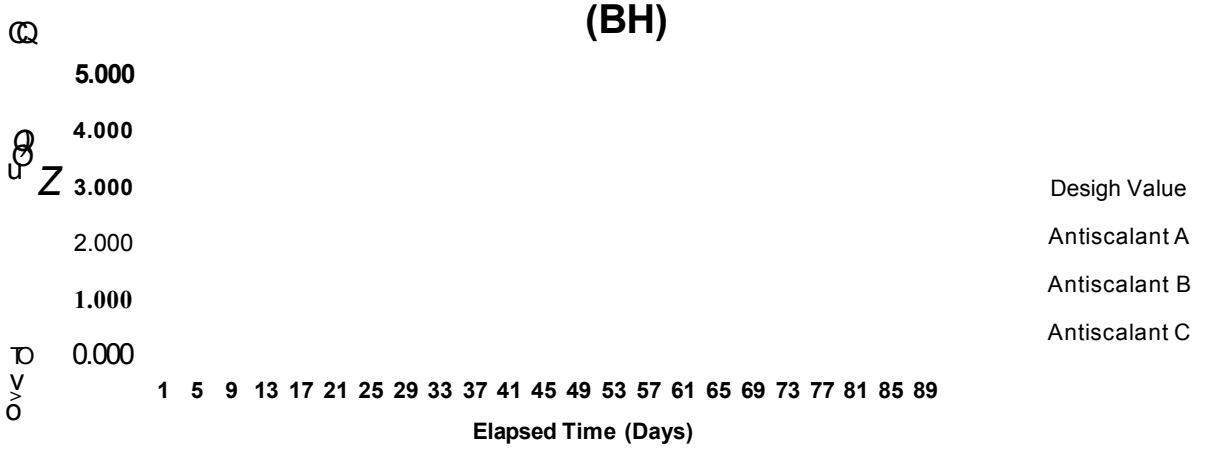


Figure 5.17: Calculated Overall Heat Transfer Coefficient (Rec. Sec.)

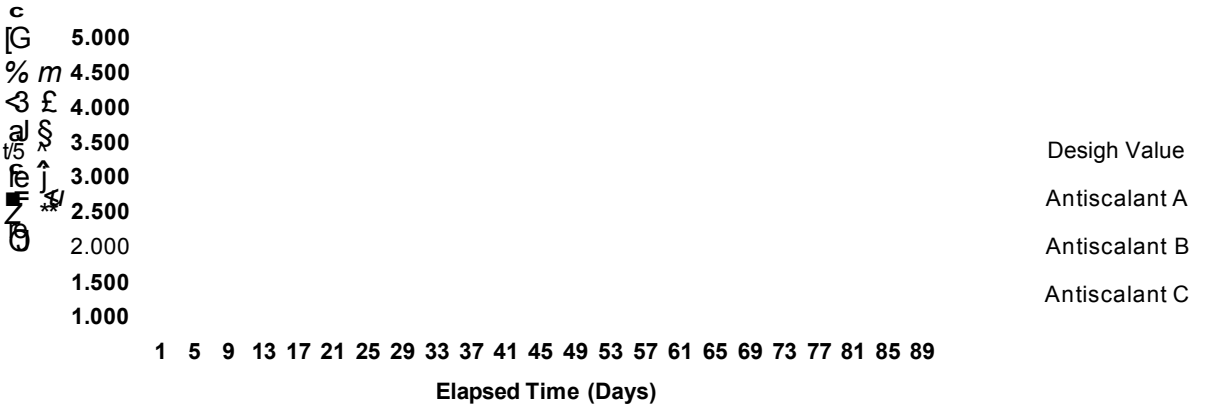
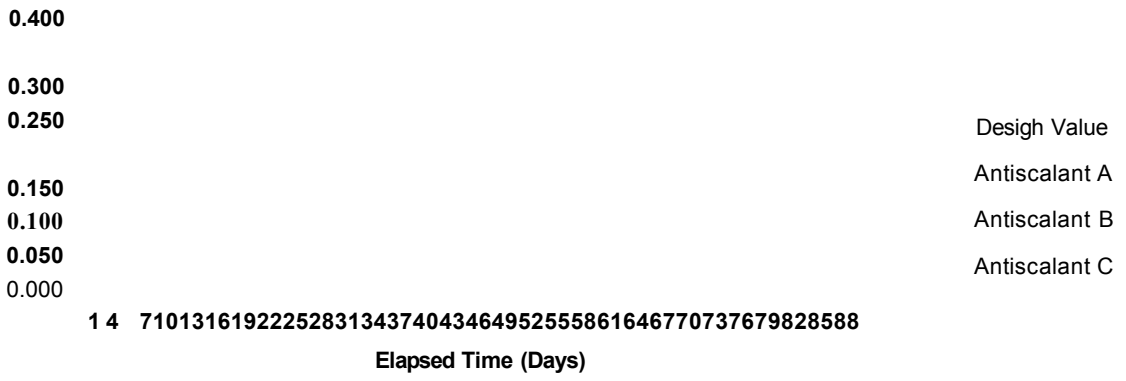


Figure 5.18: Calculated Fouling Factor, B/H.

Figure 5.19: Calculated Fouling Factor Heat Rec. Sec



5.2 Scale samples

Advantage was taken of using the X-ray fluorescence (XRF) facilities at Sheffield Hallam University to analyse the composition of samples of scale from various parts of the plant. This would be a preliminary study to see if the composition of the scale varied with location.

Scale samples were collected and analysed. The scales had different appearances with variations in density, some were multi-layered while others were smooth or rippled and some were in the form of sludge. There was also a range of different colours. In all cases, the samples were soft and non-adherent and could be easily wiped off when wet and scraped off when dry. The analysis of the following samples of the major oxides was prepared by fusing 1g of material with 10g of a flux, lithium tetraborate (Li2B40 7) at 1200 °C. The samples were also prepared as pressed pellets to determine the minor and trace elements. The analysis was determined on a Philiips PW2440 wave-length dispersive sequential X-Ray spectrometer.

At the start of a cycle the distiller will be acid cleaned and water jetted to remove scale build-up. The trials on the antiscalants were conducted over six month periods and there was no indication of excessive scale build-up. Typically the cycle would be run for one year and by that time there would be an increased build-up of scale. Although the sponge balls would help keep clear the tubes of the heat exchanger

there would be instances of blocked tubes which would result in an increase in scale. This would be due to the sponge balls not being able to clean the tube and the fact of increased residence time for brine in the tube giving rise to more deposits. There would also be a build-up of sludge on flat surfaces and this could affect the weirs and scale would also be deposited on the demister pads.

After a period of about one year the set would be shut down for a major clean that would also allow for maintenance and resetting of the weirs. The samples analysed were taken at the end of the cycle.

The samples, shown in figures 5.20 - 5.24, were taken from a distiller that had to be shut down for cleaning due to improper dosing of the scale inhibitor. This highlights the need for close monitoring of the plant.

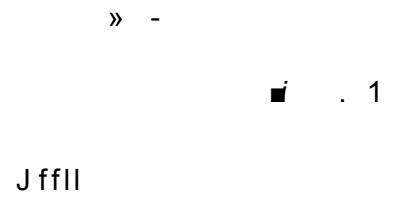


Figure 5.20: Sample no.1 from the brine heater.

Figure 5.21: Sample no. 2 from the leftside of the brine heater inlet.

Figure 5.22: Sample no.3 bottom of the brine heater inlet.

Figure 5.23: Sample no.4 from the left side of the brine heater inlet.

Figure 5.24: Sample no.5, soft scale flakes accumulated on the floor of the flash chamber and near the weir boxes.

5.2.1 Results of scale samples

Element	Unit	1	2	3	4	5
Ca	%	33.09	35.88	15.64	18.58	13.22
Mg	%	1.23	0.74	3.49	12.24	1.54
Si	%	0.796	0.323	0.37	4.02	1.11
S	%	0.595	0.466	0.295	0.286	0.849
P	%	0.113	0.0733	0.0212	0.276	0.0605
Al	%	0.339	0.0508	0.136	0.258	0.398
Zn	%	0.0085	0.0121	24.13	0.361	
Cl	%	0.552	0.123	1.37	0.35	12.45
Fe	%	0.0838	0.0768	0.0846	0.244	0.146
Sr	%	0.667	0.753	0.207	0.299	0.39
Na	%	0.835	0.31	1.89	0.108	11.37
K	%	0.056	0.0314	0.055	0.071	0.188
Ti	%	0.0099		0.0077	0.019	0.0183
V	%				0.0165	
Mn	%				0.0167	
Ni	%				0.0109	
F	%	0.043	0.037			
Br	%	0.0179	0.0137			0.0201
Cu	%	0.0117				

Table 5.4: XRF analysis (Elements)

Compound	Unit	!	i	2	3	4	5
CaO	%		46.28	50.18	21.88	25.98	18.49
MgO	%		2.04	1.23	5.79	20.29	2.55
Na20	%		1.13	0.418	2.55	0.146	15.32
Cl	%		0.552	0.123	1.37	0.350	12.45
S04	%	:	1.49	1.16	0.736	0.715	2.12
A12O3	%		0.641	0.0959	0.257	0.487	0.752
Si02	%		1.70	0.692	0.790	8.60	2.38
P_2O_5	%		0.258	0.168	0.0485	0.633	0.139
ZnO	%		0.0106	0.0151	30.04	0.450	
K20	%		0.068	0.0378	0.066	0.086	0.226
Ti02	%		0.0165		0.0129	0.0317	0.0306
V_2O_5	%					0.0294	
Fe203	%		0.120	0.110	0.121	0.348	0.209
SrO	%		0.788	0.891	0.245	0.270	0.461
MnO	%					0.0216	
NiO	%					0.0139	
F	%		0.043	0.037			
Br	%		0.0179	0.0137			0.0201
CuO	%		0.0147				
Known C02			44.80	44.80	36.00	41.50	44.80

Table 5.5: XRF analysis (Compounds

Tables 5.4 and 5.5 indicate that there is great variability between the samples. In the case of sample 5 there are high levels of sodium and chlorine which can be accounted for as the left over brine evaporating and leaving sodium chloride salt in the scale. For future analysis the samples should be washed to remove the salt so that the elements of the scale are measured.

Some variability can also be accounted for depending on which surface of the sample was measured. The initial layer of scale forming on the heat exchanger surfaces can have higher concentrations of metallic elements. This is evident in sample 3 which has a very high level of zinc compared with the other samples.

Magnesium hydroxide scale forms at slightly higher temperatures than calcium carbonate. The rate of deposition also depends on residence time and in the case of

sample 4 there are high levels of magnesium. This could be due to a blocked or restricted tube which would increase the residence time and the rate of deposition.

Chapter 6

Conclusions and Recommendations

Conclusions and Recommendations

6.1 Conclusion

The original aims and objectives of this thesis have been substantially realised.

Three antiscalants have been tested and shown to inhibit and minimise scale formation in the tubes of the distillers. Dosing rates lower than the industry standard have been investigated and shown to be effective. A monitoring and testing procedure has been instigated to provide a testing facility for future trials of new antiscalants.

The excellent scale control capability of the three different antiscalants and on-load sponge balls cleaning allowed:

- Successful operation at lower antiscalant dose rates, which would reduce the cost of chemical treatment and environmental effluents
- Possibility of increasing water production as reduced scaling will allow a number of currently operating MSF plants to increase their top brine temperatures. The antiscalants under test were able to operate effectively at high TBT (105°C).
- Improved thermal performance of the plant and hence lower running costs as scale on the heat transfer surfaces is reduced.
- Minimisation of plant maintenance as hard scale build-up would require acid cleaning and water jetting. Acid cleaning will pit the plant's structure and reduce the life of the plant.

6.2 Recommendations

Recommendations can be made based on the practical experience and knowledge gained during the study period. Some of these relate to the plant management, while others are of a technical nature and are concerned with the overall operation of the distillation unit. These recommendations are:

- To continue operation at the optimized economic dose rate of 1.5 ppm for all antiscalants. This recommendation has already been adopted by all seven MSF distillers in Kuwait.
- Further optimization tests should be trialled to investigate antiscalant dose rates lower than 1.5 ppm.
- Opportunities should be investigated to improve the ball cleaning system. This would involve choice of ball size, type of coating, number of balls and cycling times.
- Improved process control should be implemented to identify deviations from predetermined limits or functionality so that plant operation can be rectified more quickly. Process control can also ensure that the plant can be can operated at safe, reliable and efficient conditions and is not allowed to reach unmanageable and dangerous values.
- In order to evaluate the performance of any distillation unit, it is of a vital importance to ensure that all instruments are in good working order. This will enable efficient operation and a more cost effective overall utilization.
- Ensure that non-condensable gases are effectively vented from the vapour side of the flash chambers and brine heater. These gases result in poor performance of distillation plant and results in reduced distillate production, increased steam consumption and a drop in the last stage vacuum.

- There is a need for effective communication between the control room operators and the chemical staff section in order to identify and rectify any abnormalities in the operational chemical process. High distillate conductivity indication could result in the dumping of distillate production to the waste line and the need to rectify the problem by the chemical section.

- Interrelated chemical factors of desalinated water should be checked, such as: (1) Total dissolved solids of calcium, magnesium, potassium, sodium, bicarbonate, carbonate, chloride, nitrate and sulphate; high total dissolved solids are objectionable because of physiological effects, (2) Dissolved gases of oxygen and carbon dioxide; high limits will leads to corrosion effects.

6.3 Future work

There is much concern on the environmental impact of large scale desalination in the Gulf region and stringent pollution controls are being proposed by the environmental protection agencies. In order for existing plant to meet future targets it will be necessary to trial new antiscalants that have less environmental impact.

- Chemical additive suppliers are requested to develop and improve their commercial products to protect human and marine life.
- The preliminary XRF studies have shown that there is much variability in the composition of the scales in the desalination plant. A more extensive study should be undertaken to help further the knowledge of scale composition in the plant.

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Water

Introduction

Water is considered as a vital factor in the global life. Water is found in large quantities in nature. It covers about 3/4 of the earth's surface. The atmosphere contains from 1-5% of water vapour. Water also constitutes 63% of the human body, 87% of milk, 92% of watermelons and 94% of tomatoes (40).

Composition of Water

The composition by weight can be determined by passing pure hydrogen over heated copper oxide. The formed water vapour is absorbed by passing through a series of drying tubes containing KOH, $CaCl_2$ or P_2O_5. From the weight of CuO before and after the experiment, the weight of oxygen that combines can be calculated and the increase in the weight of the driers gives the weight of water (40).

Natural Waters

Rain Water: is nearly chemically pure, containing as impurities traces of dissolved gases N_2, O_2, CO_2, NH_3, HNO_2, HNO_3, NH_4NO_3 and traces of H_2SO_3 and H_2SO_4 in industrial towns. Rain carries down with its dust particles. The amount of impurity other than gases does not exceed 5 parts / million. Rain water may run on the ground as surface water and collects in streams and rivers, or may penetrate deeply into the soil as underground water and emerge as spring and mineral water or well water (40).

Sea Water: contains practically all elements dissolved and leaves on evaporation a residue of 3.6% which consists mainly of NaCl, 2.6%, and to a less extent of Mg and K salts and traces of Iodide I_2 (Iodide) (40).

River Water: this is surface water which contains a good deal of mineral matter and often a vast population of bacteria. Such water must be purified and sterilized (40).

Spring Water: was usually filtered through finely porous sand before emerging and is consequently relatively free from bacteria.

Mineral Water: this is a special type of spring water which is relatively richer in salts than ordinary spring water. Among mineral waters may be mentioned, effervescent waters rich in CO_2 and containing little alkali salts; alkaline waters containing appreciable quantities of sodium bicarbonate, chlorine and sulphate.

Sulphurous Water: this is spring water which contains H_2S and other sulphur compounds (40).

Well Water: from shallow wells is usually contaminated with bacteria. Deep wells give water which is free from bacteria as a result of filtration through porous sand. Such water is also very hard as it contains $Ca(HCO_3)_2$.

A.K. Lazlad (40) stated that: "Traditional water supply system like wells, dams, reservoirs and canals are not anymore the best solution in many areas of the world. However, new and alternative sources of water must be developed to meet the present and future needs. Oceans provide an unlimited source of water, but it is, too salty to sustain human life and farming"

Oceans are the primary source of freshwater and the ultimate sink of all water on the earth. Hence, desalination should be included as a viable option in any evaluation of water supply alternatives, not mentioning the areas where it is the only option. The future need for desalination will increase due to increasing demand for freshwater,

increasing need for water of higher quality, increasing cost of developing traditional sources of freshwater and decreasing cost of desalination technologies (41).

Physical Properties of Water

Pure water is a clear, tasteless, odourless liquid. At ordinary temperature water appears colourless in small quantities, but blue when viewed through considerable depths (40).

The specific conductivity of pure water amounts to 0.04×10^{-6} mohs, at 25°C. Owing to the presence of dissolved CO_2, the conductivity of water increases appreciably and for accurate measurements water must be freed from dissolved CO_2.

The coefficient of expansion of water varies remarkably with the temperature. Thus when water is heated or cooled at temperatures above 4°C, it follows the general rule that bodies expand when heated and contract when cooled. However, when water is cooled from 4 °C to 0 °C it expands, when it is heated from 0 °C to 4 °C it contracts. The volume of ice at 0 °C is 0.916 the volume of water at 4 °C and the volume of water at 100 °C are 1.043 times as that at 4°C. The maximum density of water is accordingly achieved at 4 °C (40).

Water exists at standard atmospheric pressure as a liquid between 0 °C and 100 °C. At 100 °C water boils and transforms into steam. The heat required to transfer 1 gram of water from the liquid state to the vapour state at its boiling point is called the latent heat of vaporization. It amounts to 2257 J are absorbed when 1 gram of ice is transformed into water at 0°, the so-called latent heat of fusion (40).

It is believed that water is associated into different types of molecules, $(H_2O)_x$ in equilibrium with the single molecules: H_2O. The association of water molecules is attributed to the fact that the hydrogen atoms do not lie symmetrically relative to the oxygen atom (40).

Chemical Properties of Water

Dissociation of water: Water is a stable compound owing to its exothermic formation from hydrogen and oxygen. It dissociates, however, at high temperatures into its elements:

$$2H_2O + 572\ kJ \rightleftharpoons 2H_2 + O_2$$

Water is electrolyzed in presence of a slight concentration of an acid or alkali yielding hydrogen at the cathode and oxygen at the anode.

Action on Metals Water interacts with the most electropositive reactive metals to form hydrogen. The more active metals react with cold water and the less active ones require steam.

$$2Na + 2H_2O\ (cold) \rightarrow 2Na^+ + 2OH^- + H_2\uparrow$$

$$3Fe + 4H_2O\ (steam) \rightarrow Fe_3O_4 + 4H_2\uparrow$$

Action on Oxides: It reacts with anhydrides (acidicoxides) to form acidsand with basic oxides to form alkalis (bases).

$$SO_2 + H_2O\ (cold) \rightarrow H_2SO_3$$

$$Na_2O + H_2O \rightarrow 2Na^+ + 2OH^-$$

Hydrolytic Reactions: Halides of non-metals react with water, forming hydrogen halides and the acid of the element:

$$PCL_3 + 3H_2O \rightarrow H_3PO_3 + 3HCl$$

Halides and other salts of metals are often hydrolyzed by water forming a basic salt together with the free acid:

$$BiCl_3 + H_2O \rightleftharpoons BiOCl + 2HCl$$

$$Bi(NO_3)_3 + H_2O \rightarrow BiONO_3 + 2HNO_3$$

Water can be considered as an acid or as a base:

$$2H_2O \rightarrow H_3O^+ + OH^-$$

$$H_2O + NH_3 \Leftrightarrow NH_4^+ + OH^-$$

Water Resources in Kuwait

The state of Kuwait lies at the northeast edge of the Arabian Peninsula at the head of the Arabian Gulf and occupies a total area of 17000km2 of mostly desert land. The hot dry climate in Kuwait results in a very low rainfall with an average of 110 mm per year. Fresh water stream do not exit. The fresh water resources are limited:

(1) Ground water
(2) Desalinated seawater and brackish ground water
(3) Treated waste water effluents.

Desalination of Seawater and Brackish Groundwater

Seawater and brackish ground water are considered as complimentary water resources in Kuwait, at present, desalination of seawater is accomplished in five distillation and power plants. The distillation units in the power stations use the MSF evaporation method (42).

Each distillation units consists of number of stages ranging between 24-26 stage and the capacity of the units is 0.023 - 0.027Mm3/d for each unit according to each station. However, the total capacity of the distillation units in the power and water distillation stations is 1.175 Mm3/d which can be raised to 1.291 Mm3/d at high temperature operation of some distiller units at Doha West and AZ-Zour South stations (42).

In order to produce fresh (potable) water for human use, distilled water collected from the seven (42) distillation plants is being mixed with underground water pumped from the brackish water wells in the blending plants. The disinfection of the blend is achieved by injecting chlorine solution, then adding the caustic soda solution to maintain the OH value of the water within the required limits according to the specification of the World Health Organization (WHO).

History of Water Supply in Kuwait

Introduction

Sixty years ago, Kuwait possessed only a coast line along the Arabian Gulf and an extended flat desert. For water people used shallow water wells and fresh water imported from the Iraqi borders by ships (Figure B.1) and then distributed to people houses in 1938 (43)(Figure B.2).

Figure B.1: Fresh water supplies from ships (42).

Like most countries possessing a coastline, the Kuwaitis depended on fishing, trading, pearl diving and boat building. At this time Kuwait was renowned for its skill in boat building; a large proportion of the dhows used throughout the Arabian Gulf countries were built in Kuwait (42).

Figure B.2: Fresh water distribution to homes (42).

Kuwait is also considered as one of the hottest areas in the world with summer temperatures averaging 35°C. In the high summer period there are daytime temperatures of 50°C in the shade and surface temperatures due to solar radiation of 84°C. Winter temperatures average 20°C but can be as low as zero Degrees Celsius (42).

In 1938, oil was discovered in Kuwait and the first oil shipments began in 1946. As a result, enough income was established to start a new era of development, modernization and industrialization which required relatively large quantities of fresh water supplies as well as electrical power supply.

The accelerated demand for both utilities was due to the substantial rise in the standard of living of the Kuwaiti population and the large increase in population due to high rates of immigration of the labour force required for the development of the country. This immigration resulted in a sudden increase in the population from 75,000 in 1937 to 400,000 in 1963. The population today is more than 2.5 million (42).

As oil production and a stable economy appeared to be permanent, the population also increased rapidly and the government realized the necessity of adopting reliable

sources of power and fresh water throughout the year. Kuwait turned to the sea to build a small power plant to supply steam to the first seawater distillation plant which was required to augment the water supply at that time, and to use the seawater for cooling the power plants (42).

From the beginning, Kuwait realized the advantages and economical savings of combining both utilities (power and water) in what is termed “Dual-Purpose Plants” and it has been the practice since the early stages of development in the early 1950’s to design and construct both electrical power generation plants and the distillation production plants side by side on the same site.

The first power and distillation plants installed in 1954-1955 each produced one million imperial gallons per day and consisted of 10 triple effect evaporators of the submerged type, similar to those originally developed as marine auxiliaries.

In 1957 the first flash evaporators, but based on multi effect design, was installed in Kuwait, and this increased the production of distillate by 2.0 MIGPD.

During 1960 the world's first multistage flash MSF distillation of the present design was installed as part of a dual purpose complex generating electrical power for distribution to consumers both domestic and industrial and steam for the distillers. The steam was extracted from the steam turbines or when there was insufficient steam available from the turbines it was obtained direct from the boilers after pressure reduction and de-superheating (42).

After the period 1964-1975, the government of Kuwait continued to increase the power generation and distilled water production plants. For power generation, Kuwait relied on conventional thermal power stations utilizing the available fuel which includes heavy oil, crude oil, and natural gas.

At present, there are seven power generation and distilled water production plants in Kuwait, Shuwaikh, Shuaiba North, Shuaiba South, Doha East, Doha West, AZ-Zour South and Sabiyah, Figure B.3. The developments of the capacities of these plants are shown in Table B.1 and Table B.2. Since 1998, the MSF units in Doha West and AZ-Zour South were modified to operate at a higher top brine temperature which raised the capacity of each unit from 6 to 7.2 MIGPD. The MSF units in Sabiyah, which are under construction, have 12.5 MIGPD each (44).

Year	Shuwaikh	Shuaiba (North)	Shuaiba (South)	Doha (East)	Doha (West)	AZ-Zour (South)	Total
1953	1	-	-	-	-	-	1
1960	6	-	-	-	-	-	6
1965	8	1	-	-	-	-	9
1968	14	9	-	-	-	-	23
1971	18	14	5	-	-	-	37
1976	18	14	30	-	-	-	62
1978	16	14	30	18	-	-	78
1980	14	14	30	42	-	-	100
1983	32	14	30	42	18	-	136
1985	32	14	30	43	96	-	215
1987	32	14	30	43	96	6	221
1991	-	-	30	42	96	48	216
1994	18	-	30	42	96	48	234
2004	19.5	-	26.4	42	110.4	115.2	313.5
2007	19.5	-	32	42	110.4	115.2	319.1
2008	19.5	-	36	42	110.4	115.2	323.1

Table B.1: Development of Distillation Plants Capacity (MIGPD) 1953-2008 (44).

Year	Number of units	Unit capacity, m^3/d (MIGPD)	Total capacity, m^3/d (MIGPD)	Type	No.	PR	Location	Status	Manufacturer
1953	10	454.6 (0.1)	4,546 (1)	Submerged tube	4	2.5-3.1	Shuwaikh	D	Westinghouse, USA
1954/1955	10	454.6 (0.1)	4,546 (1)	Submerged tube	4	2.5-3.1	Shuwaikh	D	Westinghouse, USA
1959	4	2,273 (0.5)	9,092 (2)	Flash evaporator	4	3.3	Shuwaikh	D	Westinghouse, USA
1960	2	4,546 (1)	9,092 (2)	MSF	30	5.6	Shuwaikh	D	Weirwestgarth, GB
1965	2	4,546 (1)	9,092 (2)	MSF	30	8	Shuwaikh	D	Weirwestgarth, GB
1965	3	4,546 (1)	13,638 (3)	MSF	30	7.65	Shuaiba North	D	Weirwestgarth, GB
1968	2	9,092 (2)	18,184 (4)	MSF	24	8.33	Shuwaikh	D	Westinghouse, USA
1968	2	9,092 (2)	18,184 (4)	MSF	25	8	Shuwaikh	D	IHI, Japan
1968	1	9,092 (2)	9,092 (2)	MSF	24	8	Shuaiba North	D	Westinghouse, USA
1968	2	9,092 (2)	18,184 (4)	MSF	25	8	Shuaiba North	D	IHI, Japan
1970	1	18,184 (4)	18,184 (4)	MSF	26	8	Shuwaikh	D	IHI, Japan
1971	1	22,730 (5)	22,730 (5)	MSF	25	8	Shuaiba North	D	Alsthom, France
1971/1972	4	22,730 (5)	90,920 (20)	MSF	25	8	Shuaiba South	O	Alsthom, France
1975	2	22,730 (5)	45,460 (10)	MSF	26	8	Shuaiba South	O	IHI, Japan
1978	3	27,276 (6)	81,828 (18)	MSF	26	8	Doha East	O	IHI, Japan
1979	4	27,276 (6)	109,104 (24)	MSF	26	8	Doha East	O	IHI, Japan
1982	3	27,276 (6)	81,828 (18)	MSF	24	8	Shuwaikh	O	Hitachi, Japan
1985	1	4,546 (1)	4,546 (!)	MSF	28	10	Doha East	S	Reggina, Italy
1984	4	27,276 (6)	109,104 (24)	MSF	24	8	Doha West	O	Reggina, Italy
1985	12	32,731.2 (7.2)	392,774.4 (86.4)	MSF	24	8.65	Doha West	O	Sasakura, Japan
1989	8	32,731.2 (7.2)	261,849.6 (57.6)	MSF	24	8	AZ-Zour South	O	Sasakura, Japan
1998	4	27,276 (6)	109,104 (24)	MSF	24	8.65	AZ-Zour South	O	Mitsubishi, Japan
2001/2002	4	32,731.2 (7.2)	130,924.8 (28.8)	MSF	24	8.8	AZ-Zour South	O	DHIC, Korea
2006/2007	4	56,825 (12.5)	227,300 (50)	MSF	23	9.5	Sabiya	UC	DHIC, Korea
2007	4	56,825 (12.5)	227,300 (50)	MSF	23	9.5	Sabiya	UC	DHIC, Korea

D, demolished; O, operational; S, out of service; UC, under construction

Table B.2: Progress of the MSF industry in Kuwait (45) .

Year	Description
1953	First multi-effect submerged-tube type distillation plant was installed in Shuwaikh station. The station included 10 units each with a capacity of 455m^3/d (0.1 MIGPD). These units have been demolished.
1957/58	Construction of the first flash type distillation plant in Shuwaikh station. The installation included four units each with a capacity of 2,300m^3/d (0.5 MIGPD). System features included installation of stainless steel demisters in stages to reduce carryover of brine droplet in the distillate vapour. This resulted in good control over the product salinity and maintaining it at values below 30 ppm. Installation of a strainer to remove debris from the feed seawater make-up. This reduced fouling and tube clogging in the condenser tubes in the heat recovery section and the brine heater. These units have been demolished.
1960	Installation of two MSF units in the Shuwaikh station, the world's first MSF desalination plant, each with a capacity of 4,550m^3/d (1 MIGPD). The new units had a performance ratio of 5.6. The system included a single input/single output controller on the heating steam flow rate. The controller measured the temperature of the outlet brine stream from the brine heater. This improved system operation and maintained the outlet temperature at the desired value to maintain constant production capacity. These units have been demolished.
1965	Installation of two MSF units in Shuwaikh station each with a capacity of 4,550m^3/d (1 MIGPD). The new units had a performance ratio of 8. The tube material in the first and second stages in the heat recovery section were made of 70/30 Cu/Ni alloy. The tubes in all other stages were made of aluminium brass tubes. The intake seawater temperature was controlled by mixing with part of the outlet cooling seawater. The entire system had a remote control room. Brine blow down flow rate was controlled with the brine level in the last stage. This prevented reduction in the brine level in the last stage, which may result in vapour flow across the stages. These units have been demolished.
1965	Installation of three MSF units in Shuaiba North station each with a capacity of 4,550m^3/d (1MIGPD). The new units had similar features to those installed in Shuwaikh in the same year.
1968	Installation of four MSF units in Shuwaikh station each with a capacity of 9,100m^3/d (2 MIGPD). The new units had a performance ratio of 8. The flow rate of the seawater make up stream to the last stage was controlled by the salinity of the brine recycle stream. Increase in the feed seawater make-up would result in reduction of the conversion ratio; while, its decrease would result in an increase in the brine recycle salinity. These units have been demolished.
1968	Installation of three MSF units in Shuaiba North station with a unit capacity of 9,100m^3/d (2 MIGPD). The units had the same characteristics as those installed in Shuwaikh in the same year. These units have been demolished.

1970	Installation of one MSF unit in Shuwaikh station with a capacity of 18,200m^3/d (4MIGPD). The main achievement in this new installation was to double the production capacity. These units have been demolished.
1971	Installation of one MSF unit in Shuaiba North station with a capacity of 22,700 m^3/d (5 MIGPD) and four MSF units in Shuaiba South Station. The main achievement in these new installations is the increase in the unit production capacity by 25%. The system had two methods for brine blow down, the first through the use of the brine blow down pump and the second by a discharge line installed on the brine recirculation line. Previous installation had brine blow down through the brine circulation discharge line. Installation of the pump gives better control of the system performance. Another new system feature was the use of pumps driven by electric motors instead of turbines. Stainless steel was used to construct distillate trays in stages 1-4, distillate ducts in all stages, distillate box in the last stage, and distillate pumps suction piping. The Shuaiba North unit was demolished, while the Shuaiba South units remain operational.
1975	Installation of two MSF units in Shuaiba South Station with a capacity of 22,700m^3/d (5 MIGPD). The units had the same characteristics as those installed previously in 1971 at the Shuaiba South station. These units remain operational. Refurbishment of the Shuaiba south units was started in 2005. This will extend the life of the units for an additional 10 years. The unit capacity was also increased from 5 MIGPD to 6 MIGPD. This was achieved by increasing the top brine temperature to 110°C.
1978/79	Installation of seven MSF units in Doha East Station with a capacity of 27,300 m^3/d(6 MIGPD). The main achievement was the increase in the unit capacity to 27,300 m^3/d (6 MIGPD), which remains as the standard unit capacity for all MSF installations over the next 15 years. These units remain operational.
1982	Installation of three MSF units in Shuwaikh Station with a capacity of 27,300 m^3/d(6MIGPD). These units remain operational. The units were converted in 2005 to operate at a higher top brine temperature of 110°C. This has resulted in the increase in the total production capacity of the station to 19.5 MIGPD instead of 18 MIGPD.
1983/84	Installation of four MSF units in Doha West Station with a capacity of 27,300m^3/d (6 MIGPD). For the first time in Kuwait, these units utilized the on-line ball cleaning system. These units remain operational.
1984/85	Installation of 12 MSF units with a maximum capacity of 32,700m^3/d (7.2 MIGPD). A new feature in these units is the increase in the top brine temperature to 110°C in the units with the 32,700 m^3/d capacity. This has resulted in the increase in the performance ratio to 8.65. Also, the units utilized new antiscalant polymer to withstand the high temperature. Similarly, stainless steel cladding is used to protect the plates in stages 1-6 and 24. Also, the ceiling plates in stages 1-12 were made of carbon steel and cladded with 90/10 Cu/Ni alloy.

1985	An experimental MSF unit with a capacity of 4,550m^3/d (1MIGPD) was installed in Doha East Station. The unit is designed to operate at top brine temperature of 138°C and a performance ratio of 10. The unit is out of service.
1988/89	Eight MSF units each with a capacity of 32,700m^3/d (7.2MIGPD) are installed in AZ-Zour south station. The units had a stainless steel bottom for stages 1-12, which is the roof of the lower stages. The boxes for the brine blow down outlet and the brine recirculation are made of carbon steel cladded with stainless steel. A distributed control system is used together with a back-up analogue control unit. The system included a CO_2 supply system for product re-carbonation.
1998	Four MSF units each with a capacity of 27,300m^3/d (6MIGPD) are installed in AZ-Zour south station. The units had titanium tubing for the heat rejection stages and the brine heater. The system included self cleaning filters for the feed seawater.
2001/02	Four MSF units each with a capacity of 32,700m^3/d (7.2MIGPD) are installed in AZ-Zour south station. The system had shell plates made of carbon steel cladded with stainless steel. All tube supports in evaporators and brine heater are made of stainless steel. Water boxes in the high temperature section are made of carbon steel cladded with copper nickel alloy (Cu-Ni 90/10). Shell material for vent condenser, ejector condenser, and vent gas condenser are made of 904L stainless steel instead of 316L stainless steel.
2006/07	Installation of eight MSF units in Sabiya station each with a capacity of 56,800m^3/d (12.5MIGPD). The main feature of these units is use of a single deck for all flashing stages. The unit width is limited to 20 m; however, the brine load per stage width is increased by 50% to 333.3 kg/ m s.

Table B.2 (continued): Progress of the MSF desalination industry in Kuwait, (45).

The provision of potable water and electricity necessary for construction expansion was one of the major tasks given much care and attention by the Government. In 1953, the first 1 MIGPD Distillation Plant in Shuwaikh was commissioned. It was later developed to compromise 10 (total installed capacity 32 MIGPD) distillation units.

In 1988, the Station's installed capacity was reduced to 28 MIGPD after putting out of service three distillation units (capacity 4 MIGPD) due to their low efficiency or uneconomic operation and maintenance. With the expansion of demand for fresh water and electricity, it was pressing and imperative to develop new production centres (44).

In 1965, Shuaiba North Plant commenced production of fresh water; it consists of seven 2MIGPD units, with total installed capacity of 14 MIGPD. In 1988, the station's installed capacity was reduced to 9 MIGPD after putting out of service three distillation units (capacity 5 MIGD) due to their low efficiency or uneconomic operation and maintenance. The station is now out of service due to the destruction of its equipment and buildings during the Iraqi occupation.

As for Shuaiba South Plant, the first 5 MIGPD units were commissioned in 1971. It now consists of 6 units with a total installed capacity of 30 MIGPD (44).

In 1978, three distillation units at Doha East Plant were commissioned. It now consists of 7 units with a total capacity of 42 MIGPD (44).

In 1983, three distillation units at Doha West Plant were commissioned. It now consists of 16 units with a total capacity of 110.4 MIGPD at high temperature operation (44).

Likewise, 6 distillation units at Doha and AZ-Zour Plants were commissioned in 1988; it now comprises 16 distillation units bringing up total output to 115.2 MIGPD at high temperature operation.

Figure B.3 shows the distribution of MSF plants in Kuwait and figure 4.4 the location of the Doha West Power & AZ-Zour stations which were monitored to measure plant performance.

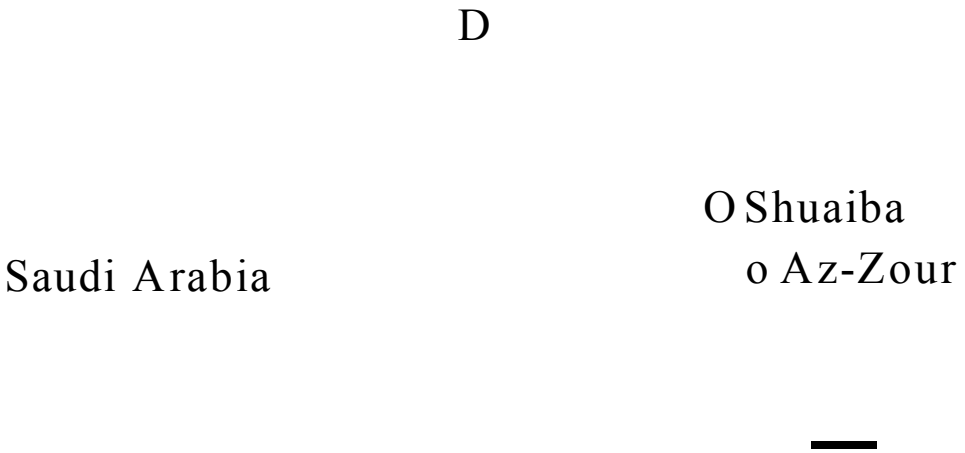

Figure B.3 Distribution of MSF distillation plants in Kuwait (31).

Figure B.4: Doha West Power & AZ-Zour Stations (39).

Description of Doha West Power Station Plant

The plant is located at Doha West Power Station which is on the Arabian Gulf near Doha harbour, as shown in Figures B.5 and B.6. The selected site for the Doha West Power Station is located at;

47° 48' longitude East

29° 22' latitude North

This is at the end of the Ashayrig peninsula, on the west side of the access road to the harbour of Doha.

The Doha West Station is the largest desalination and power plant in Kuwait. It includes 16 distillers. The total operating capacity of the Doha West units is 110.4 MIGD or 38.5% of the total desalination capacity in Kuwait, which is equal to 286.8 MIGPD. The Doha West units were commissioned in 1984/1985 by Reggiane of Italy (4 units) and Sasakura of Japan (12 units) (37). During the 1980s, all units had satisfactory operating records which met the designer's specifications production rate, product salinity and downtime.

Figure B.5: Satellite view of the Doha West Power Station (39).

Figure B.6: Seawater Intake Area of Doha west Power Station (39).

The MSF distillation plant located at AZ-Zour was commissioned in late 1988. AZ-Zour is approximately 100 km south of Kuwait city. Table B.3 shows the design information of the unit. The plant uses a cross-type MSF evaporator with re-circulating brine. The multistage condensers for the evaporator have two sections: a 21 stage heat-recovery section and a 3 stage heat rejection section. The plant consists of eight MSF-type desalting units and their common facilities. Each of the sixteen units has a daily output of 7.2MIGPD distillate, for a total of 115.2 MIGPD. Anti-scaling chemicals treatment is used to prevent scale formation inside the condenser tubes (29). Figure B.7 shows a satellite view of the plant.

Figure B.7: Satellite view of the AZ-Zour Power Station (39).

Unit	AZ-Zour South distillation unit
Manufacturer	Sasakura Engineering Co. Ltd., Mitsubishi Heavy Industries and Mitsui and Company, Ltd.
Year of commissioning	1988
Type	Cross-tube re-circulating MSF
Number of distillation units	Sixteen
Heat-input section (brine heater)	One
Number of passes	1367
Number of tubes	43.8 mm (O.D.) x 1.22 mm (average thickness) x
Tube size	18,991 mm
Heat-transfer area	Area missing
Tube material	Copper-nickel alloy 66% Cu, 30% Ni, 2% Mn, 2% Fe.
Heat-recovery section	
Number of tubes	23
Tube size	43.8 mm(OD) x 1451mm
Heat-transfer area	77,206 m^2 Copper-nickel alloy66% Cu, 30% Ni, 2%
Tube material:	Mn, 2% Fe
Stages 1 and 2	Aluminium alloy 76% Cu, 22% Zn, 2% Al
Stages 3 - 21	1588 1588
Heat-rejection section	34.2 mm (O.D.) x 9444 $m^2$66% Cu, 30% Ni, 2% Mn, 2% Fe

TableB.3: Design information for AZ-Zour South Distillation unit (27).

Distillation Station	Capacity (MIGPD)
Shuwaikh	19.5
Shuaiba	36.0
Doha East	42.0
Doha West	110.4
AZ-Zour South	115.2
Subiya **	100.0
Present Total	423.1

Table B.4: The Existing Generation Capacity of Water in Kuwait (44).

Unit	AZ-ZOUR SOUTH Distillation Units
Manufacturer	Sasakura Engineering Co. Ltd., Mitsubishi Heavy Industries and Mitsui and Company, Ltd.
Year of commissioning	1988
Type	Cross-tube re-circulating multistage flash (MSF)
Number of distillation units	8
Heat-input section (brine heater)	
Number of passes	1
Number of tubes	1367
Tube size	43.8 mm (O.D) x 1.219 mm (avg. thickness) x 18,991 mm (length)
Heat-transfer area	3,544 m^2
Tube material	Copper-nickel alloy
	66% Cu, 30% Ni, 2% Mn, 2% Fe
Heat-recovery section (H.R.S.)	
Number of passes	23
Number of tubes	1,451
Tube size	43.8 mm (O.D.) x SWG. 18 (avg.)
Heat-transfer area	77,206 m^2
Tube material:	
Stages 1 and 2	Copper-nickel alloy
	66% Cu, 30% Ni, 2% Mn, 2% Fe
Stages 3-21	Aluminium alloy
	76% Cu, 22% Zn, 2% Al
Heat-rejection section (H.J.S.)	
Number of passes	3
Number of tubes	1,588
Tube size	34.2 mm (O.D.) x SWG. 18 (avg.)
Heat-transfer area	9,444 m^2
Tube material	Copper-nickel alloy
	66% Cu, 30% Ni, 2% Mn, 2% Fe

TableB.5: Design information for AZ-ZOUR SOUTH distillation unit.

AZ-Zour South Power Generation and Water Product Station

Table B.4 shows the design information of the distillation unit used in this study. The unit, commissioned in late 1988, is located at AZ-ZOUR, approximately 100 km south of Kuwait city. The plant uses a cross-tube-type multistage flash (MSF) evaporator with re-circulating brine. The multistage condensers for the evaporator have two sections: the heat-recovery section (21 stages) and the heat-rejection section (3 stages).

The stage I distillation plant consists of eight multistage flash-type desalting units and their common facilities. Each of the eight units has a daily output of 6 .0 MIGPD distillate, for a total output of 48.0 MIGPD. Each unit is also designed to produce 7.2 MIGPD distillate during higher-temperature operation (e.g., 108-110 °C), for a total output of 57.6 MGPD.

Scale Control Additives

C.1 Seawater Pre-treatment

C.1.1 Bio-fouling

Bio-fouling in MSF plants starts with the absorption of dissolved organics to the interior surfaces of heat exchanger tubes and brine channels. While this stage of fouling would inevitably result in membrane replacement in RO systems, bio-fouling inside condenser tubes will predominantly reduce heat exchange and thereby plant efficiency (46).

Plant designers compensate for this decrease by enlarging surfaces areas. However, lasting bio-films are thought to cause corrosion (47). To control biological fouling, chlorine is added continuously at the intake. Similar to RO plants, chlorine may provide assailable nutrients when the larger organics breakdown.

C.1.2 Corrosion

Materials for construction of MSF plants are selected for their ability to withstand the harsh seawater environment and corrosion which is enhanced by high temperature, high salinity, residual chlorine and corrosive gases.

Common materials are copper-nickel alloys, used for their good thermal conductivity as well as their good corrosion resistance, which increases with nickel content. Therefore, Cu-Ni 70-30 alloys are generally used for high temperature stages and the heat rejection section.

In contrast, the cheaper Cu-Ni 90-10 is used in the low temperature stages of the heat recovery section (48), (49), where the seawater is de-aerated. Trace components further improve longevity, for example manganese and iron can be

added at 1-2% each. However, Cu-Ni alloys are not totally immune. Observed corrosion rates can remove up to 0.025 mm/annum from the surface even if oxygen has been depleted (50), and improper process control or low pH values during acid cleaning can easily increase corrosion.

Other materials can be used such as titanium for heat exchanger tubes or stainless steels for flashing chambers. Both are classified as very resistant and moreover, titanium was found to perform well in acidic environments if corrosion inhibitors are present (51). If stainless steel is used, losses of about 0.11 mm/a must be taken into account.

For stainless steel a protective layer of $Cr_2O_3$is formed on the surface (52). As stainless steels for MSF plants commonly contain 17% chromium, 12% nickel and 2.5% molybdenum (53), these ions will be released in addition to iron and may be present in the brine Table C.1 and C.2.

Item	Unit	Maximum Value
pH	-	6-8
BOD5 (5 day, 20c)	mg/L	30
COD (Dichromate)	mg/L	200
Oil / grease, Hydrocarbons	mg/L	10
Total Suspended Solids TSS	mg/L	10
Total Soluble Solids	mg/L	1500
Phosphate PO_4	mg/L	2
NH_3-N (Ammonia)	mg/L	3
NO3-N (Nitrate)	mg/L	30
Total Recoverable Phenol	mg/L	1
Fluorides F	mg/L	25
Sulfides S	mg/L	0.5
Chlorine Cl_2	mg/L	0.5
Dissolved (Oxygen DO)	mg/L	>2
Turbidity	NTU	50
Floatables	mg/L	None
Aluminium Al	mg/L	5
Arsenic As	mg/L	0.1
Barium Ba	mg/L	2
Boron B	mg/L	0.75
Beryllium Br	mg/L	0.1
Cadmium Cd	mg/L	0.01
Cyanides Cn	mg/L	0.1
Chromium Cr	mg/L	0.2
Nickel Ni	mg/L	0.2
Mercury Hg	mg/L	0.001
Cobalt Co	mg/L	0.2
Iron Fe	mg/L	5
Antimony Sb	mg/L	1.0
Copper Cu	mg/L	0.2
Manganese Mn	mg/L	0.2
Zinc Zn	mg/L	2.0
Lead Pb	mg/L	0.5
Lithium Li	mg/L	2.5
Molybdenum Mo	mg/L	0.01
Vanadium V	mg/L	0.1
Silver Ag	mg/L	0.1
Most Probable number of total Coli form	MP/N/100 ml	1000

Table C.1: Maximum Allowable values released to Kuwaiti Seawater (54).

Constituents	Units	Maximum Value
Antimony (Sb)	mg / l	0.005
Arsenic (As)	mg / l	0.010
Barium (Ba)	mg / l	0.700
Boron (B)	mg / l	0.300
Cadmium (Cd)	mg / l	0.003
Chromium (Cr)	mg / l	0.050
Copper (Cu)	mg / l	2.000
Cyanide (CN)	mg / l	0.070
Fluoride (F)	mg / l	1.500
Lead (Pb)	mg / l	0.010
Manganese (Mn)	mg / l	0.500
Mercury (Hg)	mg / l	0.001
Molybdenum (Mo)	mg / l	0.070
Nickel (Ni)	mg / l	0.020
Nitrate (NO_3)	mg / l	50.000
Nitrite (NO_2)	mg / l	3.000
Selenium (Se)	mg / l	0.010

Table C.2: Inorganic Constituents Levels of Health Significance in Potable Water (54).

More localized forms of corrosion such as pitting and crevice formation are often encountered in desalination plants, too, and increase the average heavy metal content in the brine. Local attack is of concern for stainless steel (55) as well as for copper-based material and is usually caused by residual oxygen (48) or chlorine levels (56).

C.1.3 Foaming

Seawater contains dissolved organics that accumulate in the surface layer and are responsible for white capping under wave action. In desalination plants, a surface film and foaming is not desirable as they increase the risk of salt carry-over and contamination of the distillate (46)..

To ensure the quality of the distillate and low oxygen levels for corrosion control, antifoaming agents are added to reduce surface tension in the flashing chambers and de-aeration section of the plant (46).

C.2 Scale Control Additives

C.2.1 Acid

Raw water for MSF and RO plants is commonly treated with acid as $CaCO_3$ solubility depends on pH. Precipitation becomes more likely as the brine concentration increases but can be prevented by acid addition, which keeps the pH of the solution below the pH of saturation (57). Surplus acid shifts the equilibrium to the right and thereby increases $CaCO_3$ solubility and forces CO_2 out of solution, (46):

$$CaCO_3 + HV> Ca2+ + HCO" \qquad (C.1)$$

$$HCO_3 + H+<-> CO_2 + H20 \qquad (C.2)$$

Unwanted side-effects are corrosion of pipes and brine channels or even an increased risk of sulphate scaling. However, the risk of enhanced $CaSO_4$ deposition should be minimal with regard to the natural sulphate concentration of seawater (17) given that controls help to prevent overdosing.

MSF plants face alkaline and sulphate scale formation, of which the latter is predominantly controlled by operational parameters while the first must be chemically inhibited. Thermal decomposition of HCO_3^- starts above temperatures of 50°C and carbonate ions further degrade above 80-90°C (18).

$$2HCO_3^- \rightarrow CO_2 + CO_3^{2-} + H_2O \qquad (C.3)$$

$$CO_3^{2-} + H_2O \rightarrow CO_2 + 2OH^-(2.4) \qquad (C.4)$$

Calcium carbonate is therefore likely to precipitate at relatively low temperatures when carbonate predominates in the above equilibrium Equation C.4, whereas magnesium hydroxide may form with increasing temperature. This is due to a higher availability of hydroxide ions, which are set free by thermal decomposition of CO_3^{2-} as shown in Equation C.4 and also by self-dissociation of water at higher temperature (17). However, Mubarak (23) states that only negligible amounts of hydroxyl ions are released so that $Mg(OH)_2$ crystallization is slow. Both alkaline scales are again easily controlled by HCO_3^- depletion according to Equation C.2, (58).

C.2.2 Antiscalants

Antiscalants were initially developed for use in boiling and cooling water chemistry (59). They were later implemented as scale control additives in thermal desalination plants and are nowadays commonly approved for RO systems.

Polyphosphates were the first antiscalants to be used on a commercial scale, e.g. sodium hexametaphosphate or “Shuwikh Mix” in Kuwait in 1959 (19). Polyphosphates, like mineral acids, were superseded by new polymers for the following reasons. Hydrolysis above 60°C not only lowers their efficiency but also increases the risk of calcium phosphate scaling, which is difficult to remove from heat exchange surfaces (60).

Orthophosphate, the base for polyphosphates, also serves as a nutrient and could enhance bacterial and algal growth in the plant. In contrast, new polymers like phosphonates and poly-carbonic acids are stable and highly effective for scale control in MSF plants with TBTs up to 110°C (28), (58).

Antiscalants do not react in stoichiometric proportions with scales but interact in different physical-chemical ways. Precipitation of salts is considerably delayed as divalent cations (Ca^{2+} and Mg^{2+}) are kept in solution even if solubility products are exceeded. When initial crystals are formed, further growth is stopped or modified into a fluffy structure as the inhibitor adsorbs to the crystal’s surface. It may also impart a like-charge so that particles do not further agglomerate (60). Thereby, particles are less likely to be deposited into solid scales and on-line ball cleaning systems in MSF plants continuously remove any sludge that might form. Many inhibitors also sequester other divalent ions like Fe^{2+} or Mg^{2+} by forming soluble complexes (61).

The type of treatment, i.e. acid or antiscalant or both combined the chemical nature of antiscalants and their dose levels not only differ among plants but are also only fragmentary published. Furthermore, numerous products, Table C.3 show commercially available of which chemical composition is often concealed by proprietary rights.

Poly-carboxylic acid (not further specified)	
Belgard EV2030, EV2035 and EV2050	Patel and Finan (1999)
EL-5500	Butt et al. (1997)
Poly-acrylic acid (PAA)	
Flocon 100: acrylic acid polymer	Al-Shammiri et al. (2000)
Flocon 200: acrylic acid and organophosphonate copolymer	Luu(1994)
Aqua Feed 600: acrylic acid homopolymer	Luu(1994)
Aqua Feed 1000: acrylic acid based copolymer	Luu(1994)
AA/AMPS: copolymer of acrylic acid and 2-acrylamido-2-methylpropylsulfonic acid	Gill (1999)
Hypersperse-Al: copolymer of acrylic acid and acrylamide	Luu(1994)
DP 3153: copolymer of acrylic acid and sodium phosphonate	Luu(1994)
Darex 41: polymethylacrylic acid	Butt et al. (1997)
Toflo SI-5	Luu(1994)
Pretreat 110	Luu(1994)
Degussa POC 2020	Patel and Finan (1999)
Cynamer p-70	Hasson et al.(1998)
EL-4010	Butt et al. (1997)
Polymaleic acid (PMA)	
Belgard EV: Homopolymer of maleic anhydride	Patel and Finan (1999)
Belgard EV 2000: Terpolymer of maleic acid and two aliphatic esters	Shams El Din and Mohammed (1997)
Sokalan PK 10i: polymaleic acid based copolymer	Patel and Finan (1999)
Kao KC550: polymaleic acid based copolymer	Patel and Finan (1999)
Organo-Phosphonates	
Bayhibit: 2 phosphono-1, 2, 4 – butanetri-carboxylic acid $C_7H_{11}O_9P$	Falbe and Regitz(1995)
PBTCA: 2 phosphonobutane-1,2,4-tricarboxylic acid $C_7H_{11}O_9P$	Gill(1999); Butt et al. (1997)
HEDP: 1-Hydroxyethylidene 1-1 diphosphonic acid $C_2H_8O_7P_2$	Gill (1999)
AMP: Amino tri(methylene phosphonic acid)	Gill (1999)
HMDTMP: Hexamethylene diamine tetra (methylene phosphonic acid)	Gill (1999)
PAPEMP: Polyamino polyether methylene phosphonic acid	Gill(1999)
EL-5600: polycarboxylat and phosphonat	Butt et al. (1997)
Albrivap DSB	Patel and Finan (1999)
Aquamax LT-19	Patel and Finan (1999)
Permatreat 191	Al-Shammiri et al. (2000)
Dequest 2010, 2054, 2060, 2066	Luu (1994)
Polyphosphates	
SHMP: Sodium hexametaphosphate $(NaPO_3)_6$	Gill (1999)
Calgon: Sodium hexametaphosphate $(NaPO_3)_{15\text{-}20}$ (1998)	Hasson et al.
Aqua-Mag: linear $Na_{56}H_3P_{57}O_{147}$ (1998)	Hasson et al.
Sequest-All: linear $Na_{56}H_3P_{59}O_{147}$	Luu (1994)
STP: Sodiumtriphosphate $Na_5P_3O_{10}$ (1995)	Falbe and Regitz
Nalco	Butt et al. (1997)

TABLE C.3: Commercial and non-proprietary antiscalants (58).

For example, the Saline Water Conversion Corporation (SWCC) of Saudi Arabia recommended dosing between 4.5-12.5 ppm depending on TBT between 90-110°C in 1981, which was gradually reduced to 1-2 ppm (58).

As a general rule, the required dosage will increase with the plant's TBT and conversion. General dosage recommendations for MSF have been corrected downward many times due to better antiscalant performance and awareness of chemical consumption. For example, Ministry of Electricity and Water (MEW) recommended dosing between 3.0-1.5 ppm depending on TBT between 90-110°C the levels were reduced from 3.0 ppm. Other dosage levels for RO and MSF were summarized from a literature review and are provided in Table C.4.

MSF	Dosing ppm	At outlet ppm
Antiscalings	0.8 – 1	Ahmed et al. (2000)
		Khordagui (1997)
	1 – 1.5	Imam et al. (2000)
	1 – 2	Hamed et al. (2000)
	3 – 7	Al-Ahmed and Aleem (1993)
	2 – 10	Shams El Din and Makkawi (1998)
	4.5 – 12.5	Hamed et al. (2000)
Polyphosphate	2.2+	Hamed et al. (1999)
	3 – 5++	
	2.5	Abdel-Jawad and Al-Tabtabaei (1999)
Polycarboxylic acid	1.5 – 2	Hamed et al. (1999)
BELGARD EV2000	0.53	Morton et al. (1996)
BELGARD EV2000	1 – 3	Finan et al. (1989)
EV2030, EV 2035	1 – 2	FMC (1997)
BELGARD EV2050	1 – 5	
Polymaleic acid	1.8	Hamed et al. (1999)
Polyphosphonate	1	Hamed et al. (2000)
	3	Hamed et al. (1999)
RO		
Sulphuric acid	30	Morton et al. (1996)
	82	Ebrahim et al. (1995)
	6.6 – 97	Al-Shammiri et al. (2000)
	30 – 100	Al-Ahmed and Aleem (1993)
	45 – 70	Al-Shammiri et al. (2000)
	pH 6	Ebrahim et al. (1995)
	pH 6.7	Baig and Kutbi (1998)
	pH 6 – 7	Al-Shammiri et al. (2000)
Polyphosphonate		
SHMP	2	Al-Ahmed and Aleem (1993)
	5	FilmTec (2000)
	3 – 5	Al-Shammiri et al. (2000)
	4 – 5	Al-Shammiri et al. (2000)
	5 – 10	Al-Shammiri et al. (2000)
Polyacrylic acid	2.9	Woodward Clyde Consultants (1991)
Flocon 100	1.9	Al-Shammiri et al. (2000)
Phosphonates		
Permatreat 191	1.4	Al-Shammiri et al. (2000)

Table C.4: Antiscalant concentrations in RO and MSF plants (58).

Polyphosphates result from condensation reactions of orthophosphoric acid, $(HO)_3PO$ and can be linear or cyclic in structure, as shown in Figure C.1. Cyclic polyphosphates or meta-phosphates have the total formula $(MPO_3)_n$, with M being a mono-valent cation such as sodium or hydrogen. SHMP, $(NaPO_3)_6$, belongs in this

group and the commercial product is often declared as SHMP but has the structure $(NaPO_3)_{15\text{-}20}$ (52).

SHMP forms a soluble complex with divalent cations and thereby reduces the hardness of water or complexes of heavy metal ions. Linear polyphosphates also have antiscaling properties. Their total formula is $M_{n+2}P_nO_{3n+1}$, but the actual composition can deviate slightly from this, e.g. the antiscalant Aqua-Mag $Na_{56}H_3P_{57}O_{147}$ (62). Another linear but much shorter compound is sodium tri-phosphate.

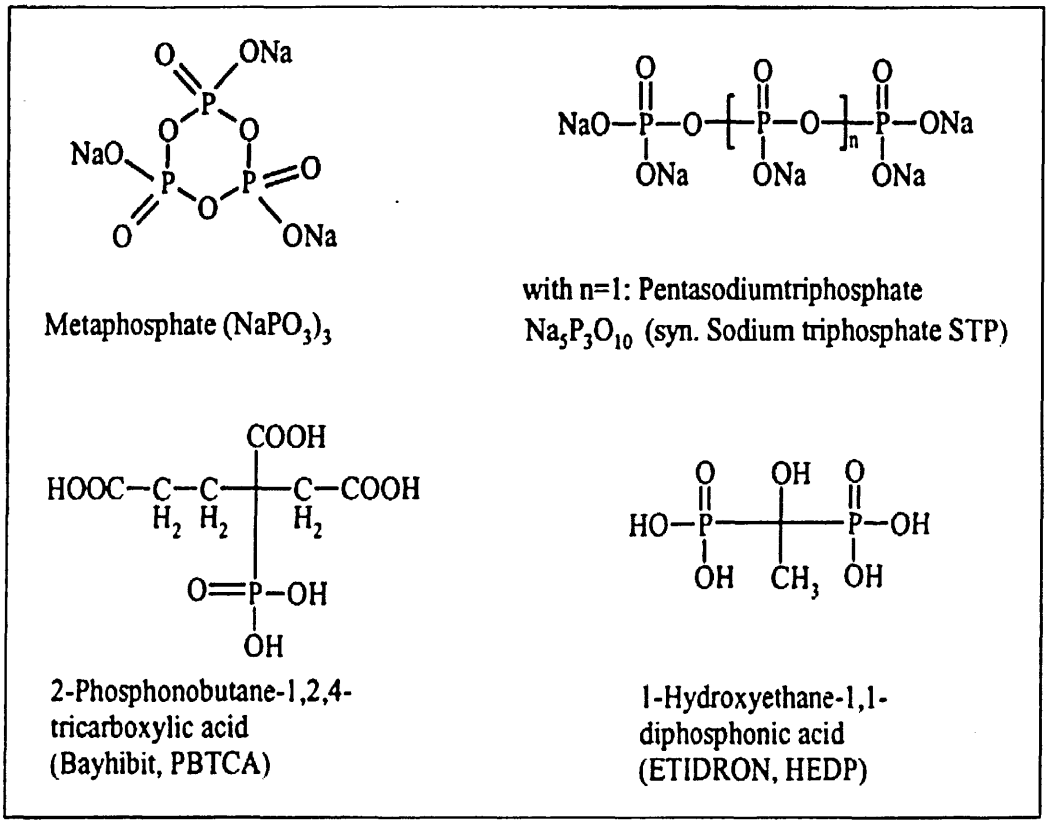

Figure C.1: Polyphosphate (top) and phosphonate antiscalants (bottom) (46).

Poly-carboxylic acids are characterized by functional –COOH groups and antiscalants in seawater desalination are usually based on polyacrylic or polymaleic structures, as shown in Figure 2.2. Polyacrylic antiscalants can be derived from homo- or copolymerization of acrylic acid or acrylic acid esters with other polymers (e.g. phosphonates). These CO^- or tri-polymers are thought to perform better in scale prevention than homo-polymers of one single monomer. The ester may be a plain alkyl residual or can contain further amino- or hydroxyl groups.

Poly-maleic antiscalants are based on maleic acid and corresponding derivates like maleic anhydride and maleates. They can again be homo-polymers such as Belgard EV, based on maleic anhydride, or copolymers with small proportions of other monomers, e.g. acrylates or even tri-polymers. An example of a copolymer is Belgard EV 2000 based on maleic anhydride with two low molar mass, unsaturated aliphatic esters (63).

Different dosing concentrations may be responsible for the contrasting properties but it has already been noted that the underlying mechanisms have not be completely understood. Agglomeration may occur if cationic poly-electrolytes and anionic antiscalants are added to the feed water (57) and in a similar way, polyvalent cations like Fe^{3+}, Al^{3+} or Cu^{2+} and anionic poly-electrolytes may form a precipitate (52).

This illustrates that each method for scale prevention has to take the overall system into account. H_2SO_4 could increase the risk for corrosion and sulphate scaling, SHMP hydrolyzes easily and thereby provides nutrients and calcium phosphate scales while anionic polymers might precipitate with cationic poly-electrolytes, calcium or dissolved cations, e.g. aluminium or iron to form a gum like deposit (57).

C.2.3 Antifoaming Agents

Antifoaming agents are common in MSF plants but are not used in RO systems, although some chemicals for RO pre-treatment such as antiscalants provide the same surface active and dispersing properties. Antifoaming agents are added to disperse the foam-causing organics in the water-air interface, which are present from excretion and degradation processes of phytoplankton in the feed water source (64).

As the additives are also organic substances, they must be carefully chosen so that they do not contribute to foam formation or increase the surface tension of seawater (52). Furthermore, another requirement is that antifoaming agents should be soluble in water over a wide temperature range: from 30°C in the low temperature stages and up to 110°C in the brine heater and they should be compatible with other pre-treatment chemicals.

Poly-glycol blends are currently in use in some plants of MEW of Kuwait. Poly-glycol usually refers to polyethylene glycol, as shown in Figure C.2 , which is linear in structure. Polypropylene glycol based substances were tested and were also found to be effective inhibitors (65).

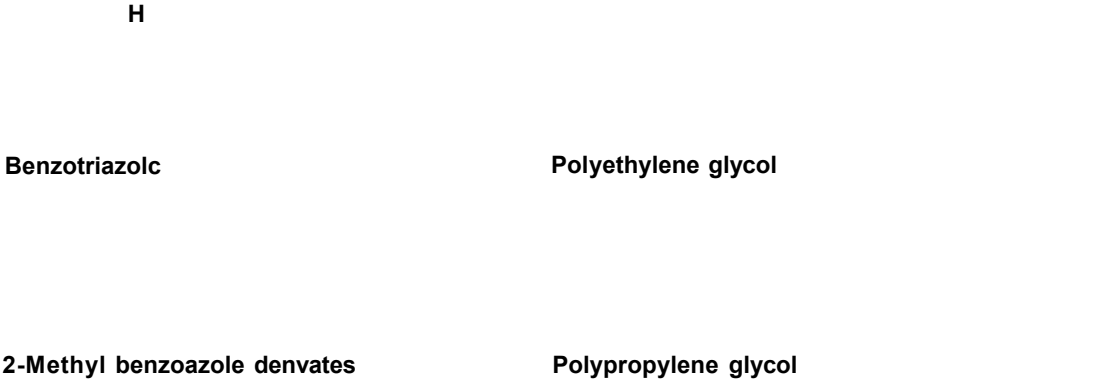

Figure C.2: Corrosion inhibitors (left, z=0: benzoxazole, z=N: benzoimidazole, z=S, Benzothiazole and z=Se: benzoselenazole) and antifoaming (right) (46).

Their chemical properties are determined by their molar masses in relation to the number of end standing hydroxide groups. They are generally soluble in organic solvents as well as in water, are very resistant to hydrolysis but sensitive to oxidizing agents such as chlorine at high temperatures.

C.2.4 Corrosion Control

To control corrosion in MSF plants, oxygen levels are brought down to approximately 100 ppb by physical de-aeration, which also depletes CO_2 in the feed stream to the heat recovery section. O_2 levels can be further reduced to about 20 ppb by reaction with oxygen scavengers (19) which is transformed into sodium sulphate.

Another preventive measure against corrosion involves addition of chemicals that form a protective film inside the tubes. For example, ferrous sulphate ($FeSO_4$) hydrolyzes and oxidizes to colloidal FeO(OH), which adheres to the surface of condenser tubes. Dosing can be carried out for one hour per day over three months

with decreasing levels from 3 to 1 ppm. In this time interval, chlorination and operation of sponge balls should be stopped as both interfere with the formation of the protective layer inside the tubes (22).

C.2.5 Polyphosphates

Sodium hexametaphosphate (SHMP, $(NaPO_3)_6$), which is still used in RO plants as antiscalant, de-polymerizes in water to sodium orthophosphate and sodium trimetaphosphate ($(NaPO_3)_3$, HSDB 2001o), and the later will be further degraded in the marine environment. Thereby, SHMP provides an essential macro-nutrient to marine primary producers.

According to Al-Shammiri et al. (2000) (66), antiscalant is used in the Al Jubail RO plant but the type of antiscalant and concentration are not specified. Based on a low dosing of 2 ppm SHMP, Al Jubail might consume 576 kg SHMP per day. In addition phosphates are used in membrane cleaning solutions and for Al Jubail, a 1% solution of SHMP amounts to 5840 kg per cleaning event.

The same amount will be necessary if sodium tri-phosphate (1%) or tri-sodium phosphate (1%) are used, and these surplus nutrients could enter the coastal water from time to time if the cleaning waste is discharged without treatment. However, no information on actual discharge from desalination plants is available. As enhanced primary production and algae mat formation near MSF plant outfalls could clearly be attributed to phosphate dosing, Shams El Din et al., 1994 (17), the same impact must be considered for RO plants in which polyphosphate antiscalants are still used.

C.2.6 Polymer antiscalants

Use of polymer antiscalants was primarily initiated for their higher efficiency at high TBTs, but Shams El Din et al. (1994) (17), reported also that the problem of algal growth totally disappeared at the outlet of MSF plants with this improved treatment. According to material safety data sheet information provided by manufacturers, polymer antiscalants have relatively high LC_{50} values, i.e. can be classified as comparatively safe for aquatic life. It seems that toxicity depends on pH, as Sokalan and neutralized Belgard EV products had considerably higher LC_{50} values than

polymaleic acid of pH 2 and the same observation was made for Flocon 100, a liquid product based on polyacrylic acid. A maximum dosage of 10 ppm will therefore be neutralized by seawater alkalinity immediately.

The amount of residual monomers might have an effect on toxicity, too. For example, about 10% of the solid content of Belgard EV can be inactive monomers (Shams El Din and Mohammed, 1997 (67)). Phosphonates like HEDP (etidronic acid) are commercially available as antiscalants, e.g. as 10-20% solution with a pH of 2.8 under the trade name Hypersperse MDC220. According to toxicological information provided by the manufacturer, organophosphonates can be classified as non-hazardous similar to other polymer antiscalants. It can be concluded that discharge levels of desalination plants are far below concentrations that could have any significant toxic or chronic effects on marine organisms.

Nevertheless, the environmental fate of polymer antiscalants is of interest and can be assessed by consulting bioaccumulation and degradation test results, as conducted for some commercial products (Wilcock and Finan, 1994 (68); FMC, 1994 (69)). For example, biodegradation of polymaleic acid is slow with 18% in 35 days (Zahn-Wellens test1), whereas polyacrylic acid was degraded almost three times faster (52% in 35 days), both yielding carbon dioxide and water.

Polymer-based antiscalants are unlikely to accumulate in marine organisms, as the tendency for accumulation increases with lipophilic properties of the chemical. Evidence for this behaviour is given by high water-solubility and direct test results that indicate low bioaccumulation factors and octanol/water partition coefficients for phosphonate-, polymaleic and polyacrylic acid-based products.

When polymer antiscalants are dispersed in the environment, they can be compared to naturally occurring humid substances. Both have high molecular masses and functional carboxyl groups, so that chemical properties might be similar, too, even though humid substances provide aromatic ring structures which do not exist in polymer antiscalants.

Furthermore, humid substances are resistant to microbial degradation and have a half lifetime of roughly 1 month, which is similar to the observed degradability of

antiscalants. As antiscalants are added as dispersing and complexion agents, these properties will not cease after discharge.

For comparison, heavy metal ions are also complexes by humid acids, which thereby prevent metal precipitation and increase mobility in the environment. Consequently, polymers might influence dissolved metal concentrations in the coastal environment in the same way as naturally occurring substances.

To conclude, polymer antiscalants are non-hazardous and comparable to naturally occurring humid material in terms of chemical properties and environmental fate. They may be preferred over polyphosphates from an environmental point of view, as they help to control nutrient input to coastal waters, but their influence on metal solubility needs further investigation.

C.3 The Development of Belgard Chemical Treatments

Ciba-Geigy's involvement in the inhibition of scale formation in seawater evaporators can be traced back to the days of World War II, when the British Admiralty placed a contract with the then Geigy Company Limited to develop a scale inhibitor for use in submerged tube evaporators (68).

Butt and Bou-Hassan (70)reported a 33% decline in Gain Output Ratio (GOR) in a 250 day test with Shuwaikh Mix, a non-proprietary polyphosphate formulation. Finan et al. (68) reported the need for frequent acid cleaning when polyphosphate was used in Qatar. It should be noted that neither of these units were fitted with on-line ball cleaning equipment.

Acid dosing was introduced in the 1960s as a means of overcoming the temperature limitations and the poor performance of polyphosphate. Acid dosing, by removing the bicarbonate from the feed water, allowed evaporators to operate at increased top temperatures, close to the calcium sulphate solubility limits.

The use of acid, however, while enhancing the thermal efficiency of evaporators and possibly reducing capital costs (lower heat transfer surface area requirement). Careful control and monitoring of the dose level was found to be essential to minimize the risks of plant corrosion or scale formation and to ensure a reasonable plant life.

BELGARD EV was originally intended for high temperature operation, but it was also found to be very effective compared with polyphosphate formulations during low temperature operation.

Many trials have been carried out to demonstrate the comparable performance of the product against polyphosphate at low temperature and against sulphuric acid at high temperature. Table C.5 gives a brief summary of four of the longer term trials carried out.

Although MSF plants in Qatar which were built in 1971 were designed for dual temperature additive operation, they continued to operate at 90/95 °C with polyphosphate up to 1975. Problems with scale and sludge formation resulted in the frequent need for acid cleaning, leading to corrosion and loss of plant life expectancy. The results from a ten month test of BELGARD EV at top brine temperatures between 105 °C and 110 °C in a unit previously operated at low temperature, are described by Diamond (71).

The dose level of BELGARD EV was 6 ppm and the unit was not fitted with on-line sponge ball cleaning. Increasing the top brine temperature of the unit resulted in an increase in production of about 20%. Diamond (71) concluded that fouling was minimal throughout the test period and that the product had proved itself as an antiscalant at 110 °C with a seawater concentration of 1.6 ppm.Table C.5 shows trials with BELGARD EV (68).

Item	Qatar	Curacao	Oman	Saudi Arabia
Plant	Ras Abu Aboud	KAE V1	Ghubrah 1	Al Khaldeah
Design TBT (°C)	90/115	105	91/113	118
Production (m^3/day)	9000/10000	8500	18000/27250	2268
Design SCA	PP/HTA	Acid	PP/Acid	Acid
SBC fitted	no	no	Yes	Yes
Date of trial	1975/1976	1978/1979	1977/1978	1983
Trial TBT (°C)	105/110	105	90/100	118
Dose of BELGARD EV, ppm	6.0	6.0	2.5/3.0	8.0
No of days' operation	300	390	520	90

TBT = Top Brine Temperature °C

SCA = Scale Control Additive

SBC = Sponge Ball Cleaning

PP = Polyphosphate

Table C.5: Trials with BELGARD EV (68).

Stewart (72) described the results in detail in 1978/1979. When operated with acid, this unit suffered from loss of thermal efficiency due to deposition of scale with a relatively high concentration of iron in the lower temperature condensers.

The unit was changed to operation with 6 ppm of BELGARD EV, again without on-line ball cleaning, and the performance was monitored for a total of 280 days operation. Stewart concluded that there was a definite advantage in favour of BELGARD EV dosing compared with acid dosing, the efficiency of the unit being maintained over a longer period. There was a reduction in corrosion rates within the unit owing to the higher recycle brine pH. During acid dosing the unit was acid

cleaned at approximately four monthly intervals, whereas with BELGARD EV, no acid washing was necessary over a period of at least 390 days operation (68).

At Al Khaldeah in Saudi Arabia, Mokhtar (73) carried out a 90 day direct comparison of plant performance operating with acid and with BELGARD EV at a top brine temperature of 118°C in a unit fitted with on-line ball cleaning. The additive was dosed at 8-10 ppm. Mokhtar (73) concluded that with regards to production, corrosion control, performance and cost savings, operation with BELGARD EV was very successful. The treatment was changed from acid dosing to BELGARD EV (68).

Between 1979 and 1989 many new MSF plants were commissioned. Some with dual scale control dosing capability (acid or additive), but the majority were designed for dual temperature operation using additive alone. All of these plants had on-line sponge ball cleaning systems included from the design stage. Many of these new and large plants were commissioned with BELGARD EV, the product being used particularly for high temperature performance and reliability tests (68).

C.3.1 BELGARD EV2000

Although many new plants were designed for dual temperature operation, the high temperature capacity, which was intended to meet future increases in demand, was not required immediately; after the initial commissioning period the plants reverted to low temperature operation. Ciba-Geigy in recognizing the need for a cost-effective additive for low temperature use, had continued research into new scale inhibitors, and in 1983 introduced BELGARD EV2000, a tri-polymer of maleic acid.

The introduction of BELGARD EV2000 followed the same pattern as BELGARD EV, with proving trials in a number of locations followed by acceptance and regular use of the product. BELGARD EV2000 is now widely used throughout the Middle East in plants operating at top brine temperatures in the range 80 to 105°C. Typical dose levels used are from 0.6 to 3.0 ppm.

Shams El Din et al. (22) compared the use of a proprietary polyphosphate formulation with BELGARD EV2000 over almost 2 years of operation of the MSF evaporators at Umm al Nar in Abu Dhabi. Evaporators of low temperature design, previously limited to a top brine temperature of 90°C, were able to operate up to 94°C resulting in an 8-10% increase in distillate production. He concluded that the use of BELGARD EV2000 gave steadier operation while maintaining plant efficiency.

C.3.2 Antiscalant Additive Characteristics of Belgard EV2000

Belgard EV2000 is an aqueous solution of a polymer based on maleic anhydride. It is corrosive and, therefore, corrosion resistant dosing equipment should be used (69).

The dose level required depended upon the type, design and operating conditions of the plant but it will normally be between 1 and 10 mg/l based on make-up feed flow-rate. The treatment solution should be dosed continuously in order to maintain the required dose level. Table C.6 lists some relevant physical properties of Belgard EV2000 (69).

Relative Density	1.18 – 1.22 at 20 °C
Physical Form	Clear Amber Liquid
Odour	Slight
Solubility In Water	Miscible
Solids Content	47 - 53% (w/w)
pH	< 1.0
Kinematic Viscosity	18 - 24 mm^2/s at 25 °C
Boiling Point	103 °C
Decomposition Temperature	> 200 °C

TABLE C.6: Belgard EV2000 Physical Properties (69).

C.3.3 Belgard EV

Belgard EV (BELEV) is a low-molar mass preparation based on maleic anhydride. It is used as an anti-scale agent for high temperature heat exchangers employing seawater as coolant.

An alternative treatment Belgard EV2000 (BELEV 2) is a tri-polymer incorporating maleic anhydride and two unspecified, low molar mass, unsaturated, aliphatic esters. It is designed for use at top temperatures between 80 and 105 °C. Both preparations are widely used as antiscalants in MSF-seawater desalination plants. BELEV 2 is also an anti-scale agent, recommended for low-temperature condensers (69).

C.3.4 Belgard EVN

Belgard EVN is a liquid treatment for the prevention of alkaline scale deposition in seawater distillation plants. Although it was developed for use in high temperature plants, it has proved effective in plants operating with a top brine temperature of 90 - 120 °C.

It was a cost-effective alternative to polyphosphate in low-temperature plants and to acid dosing in high-temperature MSF units. Belgard EV is used at low dose levels and offers advantages when compared with acid of easier handling and reduced plant corrosion; it also gives freedom from the problems associated with the use of polyphosphates. The chemical composition of Belgard EVN is that of a neutralized solution of a synthetic polymer. The active ingredient is the sodium salt of a product based on hydrolyzed polymaleic anhydride (69).Table C.7 compares Belgard EVN with other treatments (69).

Belgard EVN	Other additives	Acid dosing
Proven high temperature operation up to 120°C	Operating temperature usually limited to 90-105°C	High temperature operation up to 120°C
Minimal corrosion risk	Minimal corrosion risk	Severe corrosion risk
Minimal handling problems	Minimal handling problems	Serious handling hazard
Low dose level (1.5-12 ppm), no special storage facilities required	Dose level usually low (5-10 ppm) no special storage facilities required	High dose level (100-300 ppm), bulk storage facilities required
Accurate dosing control not required	Overdosing can cause problems	Accurate dosing control required
Liquid form, no mixing problems	Usually solids which can cause dissolving problems	Liquid form, care required on dilution
De-carbonator not required	De-carbonator not required	De-carbonator not required

Table C.7: Comparison of Belgard EV with other treatments (69).

C.3.5 Typical Properties

Table C.8 details typical properties and should not be regarded as specification limits. Because Belgard EVN is made by a batch process, there may be slight variations from these values.

Appearance	Clear amber liquid
Odour	Slight
pH	8.0 – 9.0
Relative density at 20°C	1.26 – 1.30
Dynamic viscosity at 25°C	0.130 – 0.135 kg/m s
Boiling point (760 mm Hg)	100 – 101 °C
Freezing point range	0 to -10 °C
Solubility	Miscible with seawater, brackish water and distillate in all proportions

Table C.8: Belgard EVN Physical Properties (69).

- **Thermal Stability**

The manufacturer's thermo-gravimetric analysis of the active ingredient of Belgard EVN over the temperature range 20-600°C shows that only loss of solvent water occurs up to a temperature of 300°C.

Belgard EVN was heated under pressure, have shown that the product is stable to hydrolysis under conditions which are more severe than ever encountered in operational desalination plants (69).

- **Chemical Reactivity**

Belgard EVN is readily broken down by strong oxidizing agents and has a COD value of 0.88 kg O_2/kg. Also, Belgard EVN does not contribute in any way to the foaming problems often encountered in desalination units and is compatible with commonly used antifoaming agents (69).

Belgard EVN may be used with other water treatment chemicals, such as dispersing agents and chelating or sequestering agents, including polyphosphates (69).

- **Physical Effects in Brine**

When seawater is heated, insoluble calcium and magnesium salts (calcium carbonate and magnesium hydroxide) are formed as the result of decomposition of the bicarbonate ion:

$$Mg(HCO_3)_2 \xrightarrow{\text{Heat}} Mg(OH)_2 + 2CO_2 \qquad (C.5)$$

$$Ca(HCO_3)_2 \xrightarrow{\text{Heat}} CaCO_3 + CO_2 + H_2O \qquad (C.6)$$

At temperatures below 82°C calcium carbonate predominates; at temperatures above 93°C, magnesium hydroxide becomes the major scale component (69).

Calcium sulphate scale can also be precipitated but this is not normally likely to occur below 116 °C with the concentration factors found in desalination plants, which have no more than 70000 ppm dissolved solids in the re-circulating brine. Belgard EVN will not control calcium sulphate scaling, so plants must not be allowed to operate under calcium sulphate scaling conditions (69).

- **Effect on the Environment**

Prolonged studies have shown that Belgard EVN is not readily biodegradable but it remains virtually unchanged even though the effectiveness of the biodegrading medium is altered (69).
Of equal importance, however, is the fact that Belgard EVN has been shown not to present a bioaccumulation hazard in aquatic plants or fish, or in plants grown in soil containing the product. It does not contain any phosphorus or nitrogen to act as a nutrient for micro-organisms (69).

The absolute densities of aqueous BELGARD EV and BELGARD EV2000 solutions of concentrations varying between 1 and 50 wt/vol% were determined in the temperature range 25-90°C. The linear dependence of density on solution concentration is made use of in controlling the solid content of BELGARD EV and BELGARD EV2000 preparations. 15 wt/vol % Inherent and reduced viscosities give rise to parallel straight lines over the entire concentration range (22) .

Belgard EV 2000 is an efficient anti-scale agent in MSF units. When the distillers are operated at a top temperature of 90°C, the use of BEV-2000 assured:

(a) A steady distiller performance;

(b) Improvement in the GOR value;

(c) Steadiness of the HTC;

(d) Increase in distillate production of 3.6-4.5% above design value.

In these respects Belgard EV-2000 is superior to polyphosphate. The additives evaluated in (74) are given in the Table C.9.

Abbreviation	Chemical Type	Commercial product
PMA-1	Homo-polymer of Polymaleic acid	Belgard EV
PMA-2	Tri-polymer of Polymaleic acid	Belgard EV2000
PMA-3	Copolymer of Polymaleic acid	Sokalan PM10i, Kao KC550
EM-1	Polycarboxylic acid	Belgard EV2030, Belgard EV2035
EM-2	Polycarboxylic acid	Belgard EV205 (75)0
PAA	Polyacrylate	Degussa POC 2020
P-NATE	Phosphonate	Albrivap DSB, Aquamax LT-19

Table C.9: Evaluation of chemical additives (74).

C.7 Calcium Carbonate Crystal Growth Inhibition

The results in (74) showed that PMA-1 and PMA-2 have excellent crystal growth inhibition to calcium carbonate compared to PMA-3. This suggests that not all polymaleates have the same crystal growth inhibition properties in the bulk solution phase.

Overall, the two tests showed all the additives tested can control calcium carbonate deposition but the mechanism by which they function varies. The additive that demonstrates the best calcium carbonate control is EM-1, closely followed by EM-2 (74).

C.4 Magnesium Hydroxide Control

The test was developed to measure the differences in threshold activity between additives to control magnesium hydroxide deposition. The results indicate that EM-1 and PMA-3 have excellent magnesium hydroxide control. The poly-acrylate and PMA-1 have mediocre magnesium hydroxide control. No magnesium hydroxide control was found with PMA-2, EM-2 and phosphonate. Hence PMA-2, EM-2 and the phosphonate would not control magnesium hydroxide deposition by the threshold mechanism (74) .

A new inhibitor, poly-amino polyether methylenephosphonate (PAPEMP), has been developed for scale control in water desalination. PAPEMP is very tolerant of calcium in water and is capable of controlling both calcium carbonate and calcium sulphate scales at extremely high super-saturations (76).

C.9 The Sokalan PM 10i Properties

The range of acrylic homo-polymers for detergents and cleaners, water treatment and for the industrial applications is marketed under the Sokalan trademark Table C.10 (77).

Chemical nature Sokalan PM 101 is a copolymer based on maleic acid

Properties

Physical form	Liquid
Concentration	ca. 45%
pH value	ca. 8
K-value	ca. 20
Density (DIN 51 757, 23°C)	ca. 1. 20 g/cm3
Viscosity (23°C, Brookfield, 60 rpm)	ca. 300 mPa. s

Solubility Sokalan PM 101 is miscible in all proportions with water (including sea water). It is also soluble in mixtures of water and water-soluble solvents.

K-value The K-value of a polymer is a measure of its intrinsic viscosity and thus is related to the degree of polymerization and the molar mass. The K-value of Sokalan PM 101 was determined in a 1% aqueous solution.

Application Sokalan PM 101 effectively inhibits the formation of scale due to magnesium hydroxide, calcium sulphate and calcium carbonate, which are the main sources of scale in the distillation of sea water. Sokalan PM 101 has very good performance at the elevated temperatures encountered in the MSF process.

This polymer is usually added in quantities ranging from 1-5 ppm, the exact amount depending on the quality of the sea water and the maximum temperature of the brine.

Table C.10: Properties of Sokalan PM 10i.

Appendix D

Ball Cleaning System (Taprogge)

D.1 Introduction

Seawater as used for the distillation process carries micro and macro-fouling of different types into the system. The nature of which could be inorganic (sludge, industrial waste) or, if not heavily chlorinated, organic micro-fouling (bacteria, fungi, algae, hydroids) or macro-fouling (seaweed, fish, mussels, etc) which in general cannot be controlled by chemical treatment. This type of waterborne fouling if not prevented will result in:

- A decrease in the heat transfer rate in the heat rejection section,
- Tube erosion / corrosion, as well as corrosion beneath the deposits.

Böhmer, Heinrich (78) stated that "In these components the major problem is alkaline scaling which is formed by the crystallization of supersaturated salts on heated surface, reducing the heat transfer and brine re-circulating flow. Tube fouling in these section results in an increase in steam consumption and / or product water production".

Acid dosing to destroy the bicarbonate ions, which are a cause of scale, can be an effective countermeasure but the practice adds a risk for enhanced plant corrosion. Treatment with chemical additives prevents the formation of hard scales; however the additives may cause sludge to be deposited in the tubes, again reducing heat transfer.

Scaling can occur in the flash chambers of the heat recovery section, and in addition, corrosion products, both of which can flake off the chamber walls, and be carried by the re-circulating, brine. The scale and corrosion chips can clog evaporator and brine heater tubes, and may also cause problems with the On-Load tube cleaning system. For this reason, debris filters are now installed in the brine re-circulating pipeline.

The combined use of chemical additives and an on-load tube cleaning system has proven to be the most cost-effective means to avoid tube fouling. Operating cost reductions result from a possible reduced cost for chemical additives due to lower chemical dosing rates, reduced steam consumption, and increased production.

General Solutions to avoid Tube Fouling

- Increased water temperature (heat kills organic components).
- Alteration of seawater salinity.
- Reversed flow direction of cooling water in the condensers (backwash).
- Toxic paint on areas in contact with cooling water.
- Increased cooling water velocity in pipes and channels up to approximately 3 m/s.
- Cooling water chlorination.
- Except for chlorination, the above solutions were scarcely applied in cooling water circuits.
- The reasons are cost and space-intensive equipment, high energy input, maintenance problems (e.g. toxic paint) and the toxic and pollutant effects on marine life.
- In addition these methods are only partial solutions, e.g. chemical and thermal methods may reduce marine life, but do not influence inorganic debris.
- Furthermore, in a number of countries, depending on the power station location, chlorination of the cooling water is not allowed or partly restricted due to environmental protection regulations.

For best protection of MSF units against macro / micro-fouling and scaling, a complete system, with monitoring is the Taprogge System (Ball cleaning system), as shown in Figure D.1, and herein described.

Figure D.1: Types of Taprogge System (Ball Cleaning System) (79).

Bohmer, Heinrich (78) stated that "Erosion / corrosion are a common cause of failure in seawater cooled heat exchangers with copper base alloy tube, such as in the heat rejection section of MSF units. Failure can usually be attributed to debris, such as shells, stones or twigs, which become lodged in the bore of the tubes. This is also a common failure mode in power plant steam condensers".

The mechanism of failure can be attributed to a critical shear stress corresponding to a critical cooling water velocity being exceeded at the cooling water / tube surface interface. The usual protective oxide film is lacking and attack of the tube material can therefore proceed virtually unimpeded. Tube penetration in a few thousand hours is common (78).

Such attack is particularly troublesome with aluminium brass tubes. The critical shear stress is not all that much higher than the shear stress found at normal cooling water velocities. A small reduction in the free cross section area is then often sufficient to produce protective film breakdown and localized attack. It is not restricted to this alloy.

Self cleaning debris filters arranged at the inlet of the MSF heat rejection section are an effective solution (78).

Bohmer, Heinrich (78) stated that “On-load tube cleaning systems using sponge rubber balls as a mechanical means to remove deposits from tubs are well established as a method to maintain the cleanliness of heat exchanger tubes. The sponge rubber balls, shown in Figure D.2 which is slightly larger than the inner tube diameter pass through the tubes periodically depending on the fouling tendency. The number of balls in circulation depends on required frequencies and the time for one cycle “.

Figure D.2: Sponge rubber ball pass through the tube (79).

Although scaling problems are not often encountered in temperature zones pertaining in the heat rejection section of a MSF unit, fouling can still occur. The nature of the deposits will be similar to those found in power station condensers (biological fouling, suspended solids fouling, corrosion fouling) (78).

On-Load ball cleaning systems permit much lower fouling factors, and when considered in the design phase, demonstrate considerable saving in capital investment.

Ball cleaning will not only remove sludge from tubes but will also reduce the possibility of scale formation. It has been found that the dosing rates can be less than one third if a tube cleaning system is in operation. By the combined use of proper chemical treatment and ball cleaning operation, down times for acid cleaning are said to be completely avoided.

The necessity for using On-Load tube cleaning systems at any brine temperature and in conjunction with any chemical treatment has now been established.

Before brine enters the flash chamber the balls will be separated from the flow by means of a strainer section, and then re-injected into the heat recovery section (78).

The design of the tube cleaning system provides for a short cycle whereby the injection into stages of higher temperature may be chosen as they could require more frequent ball passages than the stages of lower brine temperature. The cleaning balls are designed to meet different operation conditions and will be selected according to the temperature of circulating water, the nature of deposition (hard, soft, slimy), and the existing pressure, loss between the stages which is the driving force for the balls (78).

The Monitoring System

The first installation of an On-Load tube cleaning system for an MSF unit by the TAPROGGE Company was in the Bahamas, in 1965. Since then tube cleaning systems have been applied for many of the major and land based MSF plants. During the intervening period the design of tube cleaning systems has passed through several generations. The current system looks significantly different, and a large amount of field and laboratory test experience has been gained for recommended operational regimes:

- Ball charges of 25% of the number of tubes per stage
- Increased cleaning cycles (up to 48 per day).
- Mixtures of cleaning balls of different specific weights to ensure optimum distribution in all stages.
- Two separate cleaning loops for heat recovery / heat input section serviced by one tube cleaning system.

In spite of remarkable progress over recent years there was still one painful gap left, namely an accurate monitoring system for the tube cleaning which instantly and directly indicates any undesired tube fouling.

Advantages of TAPROGGE Tube Cleaning System

The TAPROGGE systems economically solve:

- The problems of tube clogging by organic and inorganic coarse particles
- Tube fouling
- Low, favourable use of cooling water in circuits with respect to environment
- Reduce application of chemicals, e.g. chlorine
- Only allow a good film formation in the tubes
- Increased availability of power plants and desalination plant
- Eliminate manual cleaning of condensers and heat exchangers
- Avoid physical harm to personnel
- Capital return within approximately 0.5 to 2 years.

Appendix E

Chemical Analysis of MSF Plant Procedure

Chemical analysis tests of the collected samples were done during the trials are as following:

Conductivity Measurement Procedure

1) Place the sample in the beaker; wash the Conductivity meter probe with the sample itself for a few seconds.
2) Throw the sample after the washing step.
3) Place a new sample in the beaker; place the beaker under the Conductivity meter probe.
4) Wait a few seconds until the device give a stable reading.
5) Take the reading appears on the device screen.
6) Repeat the above steps for all samples.

pH Procedure

The pH of a sample of water is a measure of the concentration of hydrogen ions. The term pH was derived from the manner in which the hydrogen ion concentration is calculated, it is the negative logarithm of the hydrogen ion (H^+) concentration. The pH scale ranges from 0 to 14. A pH of 7 is considered to be neutral. Substances with pH of less than 7 are acidic; substances with pH greater than 7 are basic.

The pH of water determines the solubility (amount that can be dissolved in the water) and biological availability (amount that can be utilized by aquatic life) of chemical constituents such as nutrients (phosphorus, nitrogen, and carbon) and heavy metals (lead, copper, cadmium, etc.). For example, in addition to affecting how much and what form of phosphorus is most abundant in the water, pH may also determine whether aquatic life can use it. In the case of heavy metals, the degree to which they

are soluble determines their toxicity. Metals tend to be more toxic at lower pH because they are more soluble.

pH Meter

A pH meter is a complicated device. It consists of two basic parts: a pH electrode and the electronics of the meter itself.

Combination pH electrodes used in both of these systems actually contain two different electrodes inside of them. One is a reference electrode that does not change voltage, but just sets a standard voltage level to which the sensing electrode is compared. The second electrode is sensitive to pH. It is the voltage difference between these two electrodes that the meter reads and converts into pH.

pH Meter Procedure

1) Place the sample in the beaker, wash the pH meter probe with the sample itself for a few seconds.
2) Throw the sample after the washing step.
3) Place a new sample in the beaker; place the beaker under the pH meter probe.
4) Wait a few seconds until the device take a reading.
5) Take the reading that appears on the device screen.
6) Repeat the above steps for all samples.

Residual Chlorine (By Lovibond Comparator)

1) Measure the residual chlorine at the intake area by using comparator-
   N, N–diethyle-P-Phenylendiamine and add 1 (DPD1) tablet.
2) Dissolve DPD1 tablet in the sample, colour will changes.
3) Using comparator disc adjust to get matching reading.

* N, N-diethyle-P-Phenylendiamine LovoBond-balin-DPD Tablets for Residual Chlorine

Chloride Ratio (By Titration method)

1) Take using volumetric pipette grade A of size 25 ml sea water sample and dilute with De-mineralized water in a measuring flask of size 1 litre, and mix well.
2) Take the sample by using volumetric pipette of size 10 ml grade A accurately in a conical flask 250 ml from the diluted solution in step 1.
3) Add 3 drops of K_2CrO_4 (Potassium Chromate) indicator then mix it well.
4) Start the titration with (0.02N) $AgNO_3$ (Silver Nitrate) solution till the end point.
5) End point is when all Cl^- ions in the sample are precipitated as, white AgCl (Silver Chloride) the first drop of $AgNO_3$ solution will react with the indicator K_2CrO_4 to give the red Ag_2CrO_4 (Silver Chromate) which is the end point.

Appendix F

Instrumentation Photos

Figure F.1: Schematic board of the MSF technique distiller.

Figure F.2: Digital devices of distillate and return condensate flow rates.

Figure F.3: Digital devices of distillate conductivity and spray water flow rates.

Figure F.4: Indicators of 1st and 21st stage condensate vapour temperatures.

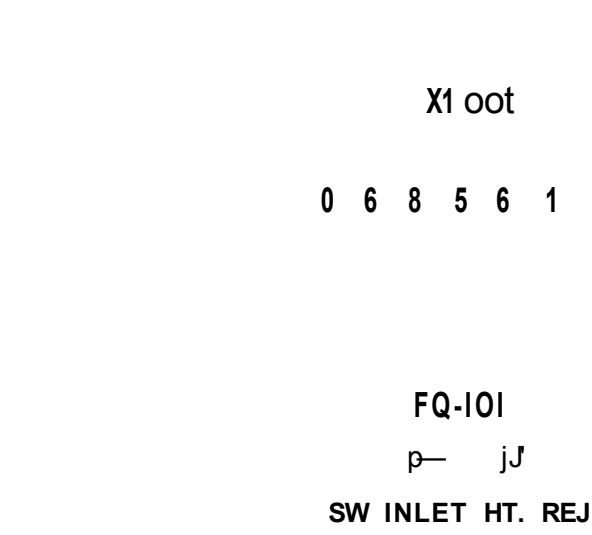

Figure F.5: Recorder of seawater inlet flow rate.

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Figure F.6: General view of the control panel.

VAP SPACt OF B/H

Figure F.7: Indicators for brine heater shell pressure.

FQ-702

® SPRAY WATER TO
DESUPER HEATER

Figure F.8: Recorder of spray water to de-super-heater flow rate.

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4 6 0 8 6 5

FQ—701

^rOTAL CONDENSATE FROM'
BRINE HEATER .

Figure F.9: Recorder of condensate flow rate.

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m

FQ-612

v STEAM TO
DESUPER HEATER

Figure F.10: Recorder of steam flow rate to de-super-heater.

FQ-6II

EJECTOR STEAM

Figure F.11: Recorder of ejector steam flow rate.

FQ-60I

© ^JLI ®

LP STEAM TO BRINE HEATER

Figure F.12: Recorder of LP steam flow rate to de-super-heater.

Figure F.13: Recorder of sodium sulphite flow rate.

FQ–501

® POLY PHOSPHATE TO DEAERATOR

Figure F.14: Recorder of poly phosphate flow rate.

FQ-412

DISTILLATE WATER PRODUCT

Figure F.15: Recorder of distillate flow rate.

FQ-321

BRINE BLOW DOWN

Figure F.16: Recorder of brine down flow rate.

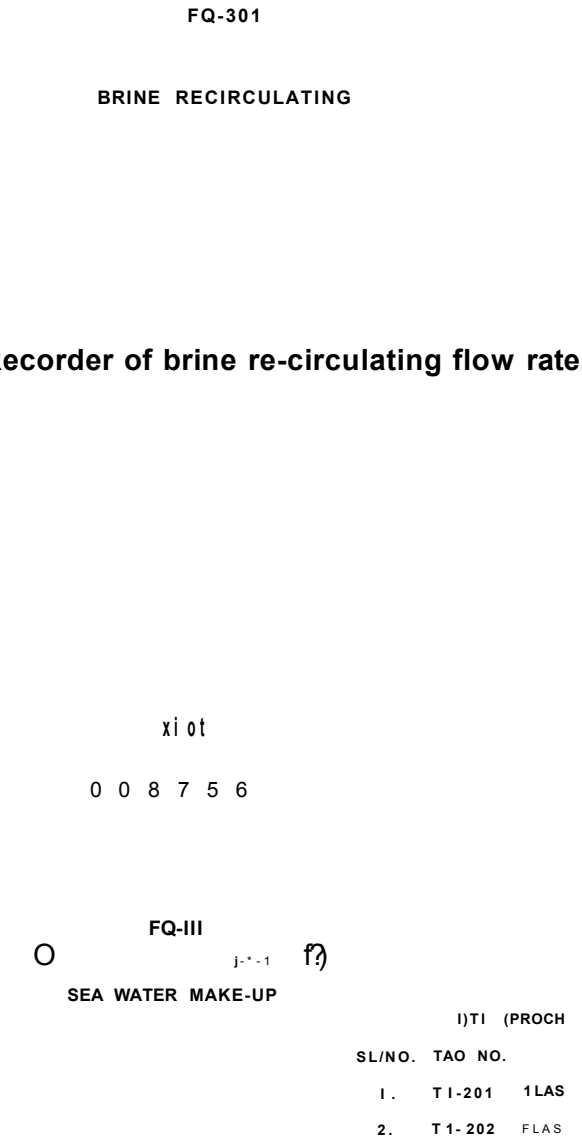

Figure F.17: Recorder of brine re-circulating flow rate.

Figure F.18: Recorder of seawater make-up flow rate.

Flow Transmitter

The Transmitting unit employing an electrically balanced beam system converts the input displacement of the meter body into a current signal of 4 to 20mA DC proportional to the input applied.

The transmitter will operate with any receiver having a load impedance (including line resistance) from 600Ω (24V) to 1400 Ω (42 V) (37).

No load adjustment is necessary.

Dead band: 0.1% of Full Scale (Sensitivity is 0.05 % of Full Scale)
Accuracy: 0.5 % Full Scale

Pressure Transmitter

This transmitter consists of an electrical beam which converts measured process pressure into a proportional electrical signal of 4 to 20 mA DC. This transmitter will operate with any receiver having load impedance up to 1,400 ohms and no load adjustment is necessary (37).

Dead band: 0.1% of full scale or less
Accuracy: 0.5 % Full Scale
Hysterics: 0.5 % Full scale or less

Pressure Transmitter (Absolute Type)

This transmitter consists of electrical beam which converts measured process pressure into a proportional electrical signal of 4 to 20 mA DC. This transmitter will operate with any receiver having load impedance up to 1,400 ohms and no load adjustment is necessary. Measuring range is continuously adjustable 1:10 of span by span adjustment. The suitable pressure range for measurement or control can be selected. The meter body is equipped with a hermetically sealed-up vacuum chamber by welding construction that assures from 0 kg/cm^2 absolute (37).

Detector (for measuring conductivity)

In solution conductivity measurement, the cell constant K, a factor which depends on the electrode configuration, is related with solution resistance X and solution conductivity K and also when a solution conductivity is measured using the electrode of cell constant K with the temperature compensator T and the source voltage Vi of the oscillator, the output voltage of the OP amplifier Vo. The output in proportion to the solution conductivity is thus obtained. Automatic temperature compensation can be established by making the characteristics of T equal to those of X. The OP amplifier output V0 provides the meter indication, and the transmission output of 4 ~ 20 mA DC is available through the converter (37).

Square root Extractor Module

This module accepts a 1 to 5 V DC input signal and converts 1 to 5 V DC output signal and thus the output signal are proportional to the input signal (37).

Resistive Thermal Device (RTD) to Voltage Converter

This module converts an input signal from a single 3-wire resistance temperature bulb (Pt) into a 1 to 5 V DC or a 4 - 20 mA output. This module provides isolation between input and output circuits (37).

'AL-HAMD-LELLAH'

'THANKS TO GOD'